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# **Nitrification-Denitrification of Wastewater Using a Single Sludge System Volume II**

## **Research Report No. 96**



**Research Program for the Abatement of Municipal Pollution  
under Provisions of the Canada-Ontario Agreement  
on Great Lakes Water Quality**

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NITRIFICATION-DENITRIFICATION OF WASTEWATER  
USING A SINGLE-SLUDGE SYSTEM  
VOLUME II

by

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RESEARCH PROGRAM FOR THE ABATEMENT  
OF MUNICIPAL POLLUTION WITHIN THE  
PROVISIONS OF THE CANADA-ONTARIO  
AGREEMENT ON GREAT LAKES WATER QUALITY

Project No. 71-1-20

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Cat. No. En 43-11/96  
ISBN 0-662-10711-X  
BEAUREGARD PRESS LIMITED

## ABSTRACT

This report describes a full-scale evaluation of the split-return mode for nitrification-denitrification of a combined domestic/industrial wastewater. The investigation, which took place from July 1974 to December 1975 at the Ontario Ministry of the Environment's Ontario Experimental Facility at Brampton, followed pilot-studies at Kleinburg, Ontario.

The evaluation investigated the effect of ferric chloride addition on phosphorous removal; the effect of solids retention times and sludge ages on nitrogen reactions at various temperatures; hydrogen peroxide addition and the use of a screen centrifuge on mixed liquor conditioning; as well as the suitability of the system for solids control and for maintenance of consistent nitrogen removal under different temperature conditions.

It was concluded that the single-sludge, split-return system gave excellent nitrogen removal with minimal methanol application. Control of solids in both aeration and denitrification basins was easily accomplished by adjusting the return-split. On-site monitoring of the process was also easily accomplished. The lack of a clearly defined temperature effect on nitrogen metabolism was attributed to the long sludge retention time inherent in the system.

## RÉSUMÉ

Ce résumé fait le point sur une expérience tentée de juillet 1974 à décembre 1975. Il s'agit d'une évaluation, en grandeur réelle, des possibilités offertes par une recirculation divisée pour nitrifier et dénitrifier un mélange d'eaux usées industrielles et domestiques. Il est à noter que cette expérience a eu lieu à l'installation expérimentale de Brampton, propriété du ministère de l'Environnement de l'Ontario, et qu'elle faisait suite aux études-pilotes réalisées à Kleinburg, en Ontario.

L'objet d'une telle évaluation était de déterminer l'effet du chlorure ferrique sur la déphosphatation; l'effet du temps de retenue des matières en suspension (MES) et du vieillissement des boues sur les réactions de l'azote, et ce à températures variables; l'effet du peroxyde d'hydrogène et de l'action d'une centrifugeuse à tamis sur le traitement de la liqueur mixte; et la capacité de l'installation à maintenir la teneur en MES et à éliminer l'azote à différentes températures, de façon constante.

Il appert que la recirculation divisée, appliquée à un type unique de boue, élimine très bien l'azote et n'ajoute que très peu de méthanol. En réglant de débit de la recirculation, il fut facile de maintenir la teneur en MES dans les bassins d'aération et de dénitrification. Le contrôle sur place du procédé n'a pas causé non plus de difficulté. Il a été impossible d'établir clairement l'effet de la température sur le métabolisme de l'azote. On a cru que cela était dû à la longue période de retenue des boues que nécessite ce genre de procédé.

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## CONCLUSIONS

The following conclusions were derived from full-scale nitrification-denitrification studies at the Ontario Experimental Facility:

1. The single-sludge, split-return system proved to be a manageable process, and operator attention was little more than that required by a conventional activated sludge process.
2. Under most conditions good nitrogen removal, ranging between 87 and 90%, was achieved at wastewater temperatures of 5 to 22°C.
3. Methanol demand for nitrate removal was generally less than theoretical. Methanol requirements were reduced on raising denitrification solids and lowering the nitrate-to-solids ratio.
4. Low levels of effluent BOD<sub>5</sub> were attainable at secondary clarifier overflow rates of less than 24 m<sup>3</sup>/m<sup>2</sup>/d (50 Igpd/ft<sup>2</sup>).
5. Application of ferric chloride reduced SVI and improved secondary sedimentation, as well as achieving a high level of phosphorus removal; little effect was observed, however, on nitrification-denitrification reactions.
6. Approximately 4.7 kg of calcium alkalinity per kg soluble-N was returned to the system by denitrification reactions. The net alkalinity loss through the system was less than the assumed theoretical value.
7. The single-sludge, split-return system had a solids yield similar to that of an extended aeration process.

### RECOMMENDATIONS

1. The single-sludge, split-return system should be considered as one of the alternatives for biological nitrogen removal, pending an evaluation of economic considerations.
2. Further long-term studies with this system using chemical addition for phosphorus removal are suggested to identify:
  - (a) required chemical dosage,
  - (b) resultant sludge yield,
  - (c) effects of SVI,
  - (d) system alkalinity balance.
3. Studies involving the use of a screen-centrifuge concentrator as a thickener for denitrification mixed liquor are also recommended.

During the past decade, considerable effort has been directed towards the removal of nitrogen from sewage effluent by biological means. This has been a result of the international community becoming aware of the need for carbon, nitrogen and phosphorus reduction in sewage effluents to protect the ecology of the often-sensitive receiving waters.

Extensive laboratory investigations have contributed much to the knowledge of nitrogen reactions within activated sludge, and parallels can be seen between biochemical exchanges that occur in soils and surface waters and the biological sewage treatment process [1]. Since the biological process appears to offer the most economical method of treating sewage, particularly on a large scale, many authorities have chosen this form of treatment for the removal of nitrogen. Full-scale evaluations, using existing or slightly altered conventional activated sludge plants, have been conducted to study nitrification-denitrification efficiencies under varying climatic conditions. Some of these studies are discussed in the following sections.

#### 1.1 Review of Full-Scale Biological Nitrogen Removal Processes

##### 1.1.1 Wuhrmann's studies

Wuhrmann [2] concluded in 1963, from studies on the effects of oxygen on sewage purification, that regardless of oxygen concentration in aeration mixed liquor, full conversion of ammonia to nitrate (nitrification) would not take place unless adequate hydraulic detention was offered in the aeration basin. Furthermore, nitrification did proceed well in experiments where the aeration basin dissolved oxygen (DO) was only 1 mg/L under conditions of low organic loadings and high sludge concentrations (about 3000 mg/L). Some attention was also focused on the possibility that individual aeration floc size and/or shape may have some bearing on floc internal DO levels, which in turn would affect nitrogen conversion efficiencies.

Periods of anaerobiosis of aeration mixed liquor (ML) of up to five hours showed little effect on subsequent endogenous bacterial respiration or substrate utilization when the system was returned to the aerobic state. Moreover, during the anoxic stage, nitrate reduction to

nitrogen gas (denitrification) took place at a high rate. From these experiments, Wuhrmann concluded that the efficiency of the aeration ML to metabolize organic materials is not deterred after a period of anoxia.

Continuing studies by Wuhrmann [3] involved using a  $60.5 \text{ m}^3/\text{d}$  (13 300 Igpd) activated sludge plant for nitrification-denitrification, followed by clarification and clarifier sludge return to the nitrification basin influent (Figure 1a). Total nitrogen removals obtained with this system ranged from 35% in winter to 90% in the summer months. These data were obtained without the use of an external carbon addition to the denitrification reaction basin and during a one-year observation period. The hydraulic detention times (normally two hours) and solids concentration were raised during periods of low wastewater temperature, but nitrification and denitrification still did not proceed fully. A minimum of four days' sludge age was required to maintain nitrification at  $10\text{--}12^\circ\text{C}$  sewage temperature.

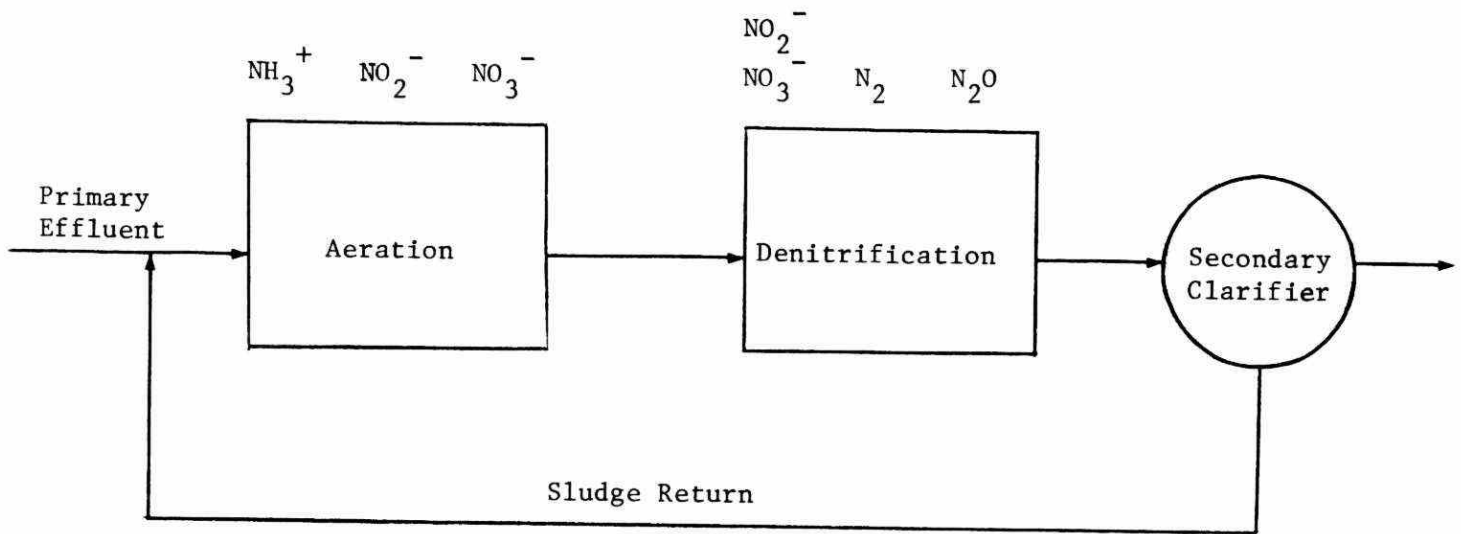
#### 1.1.2 Studies by Bishop and co-workers

Bishop *et al.* [4] have reported a process variation to the single-sludge system. Two basins were operated alternately, on a 30-minute cycle, as aerobic and anoxic reaction basins for the nitrification-denitrification process. These studies were performed over nine months on a plant designed for  $189 \text{ m}^3/\text{d}$  (50 000 Igpd) sewage flow (Figure 1b).

Addition of methanol to the process mixed liquor in the chamber between the last reactor basin and the secondary clarifier was an option when residual nitrate concentrations escaping the main basins were too high. A conventional sludge recycle from the secondary clarifier was used as was, on occasion, an internal aeration recycle.

Ferric chloride was added, at a dosage of  $45 \text{ mg/L}$ , to the primary clarifier flow at one stage of the study to remove phosphorous. Unfortunately, this treatment lowered the chemical oxygen demand/total Kjeldahl nitrogen ratio (COD:TKN), resulting in a decrease of 20% in nitrogen removal efficiency. Aluminum sulphate at a dosage of  $20 \text{ mg/L}$  was added prior to tertiary filtration for effective phosphorus removal during two months of the project.

a. Wuhrmann [3]



b. Bishop [4]

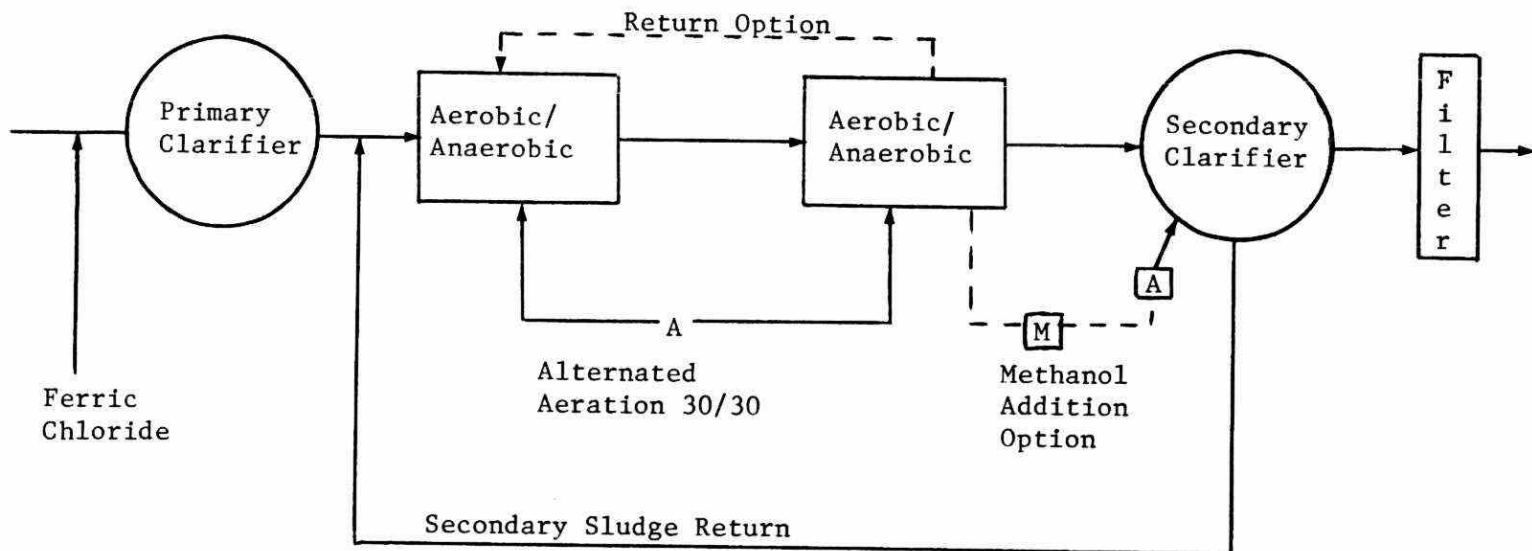
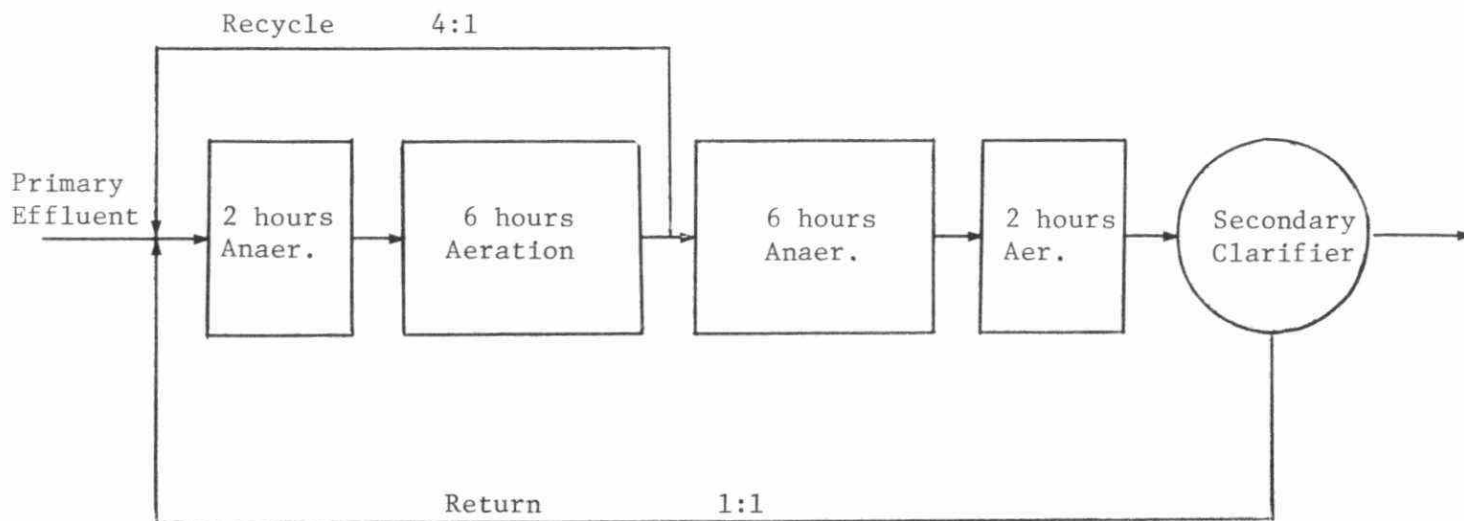


FIGURE 1. NITRIFICATION-DENITRIFICATION SYSTEMS [3,4,5,6]

c. Bardenpho process [5]



d. Nicholls [6]

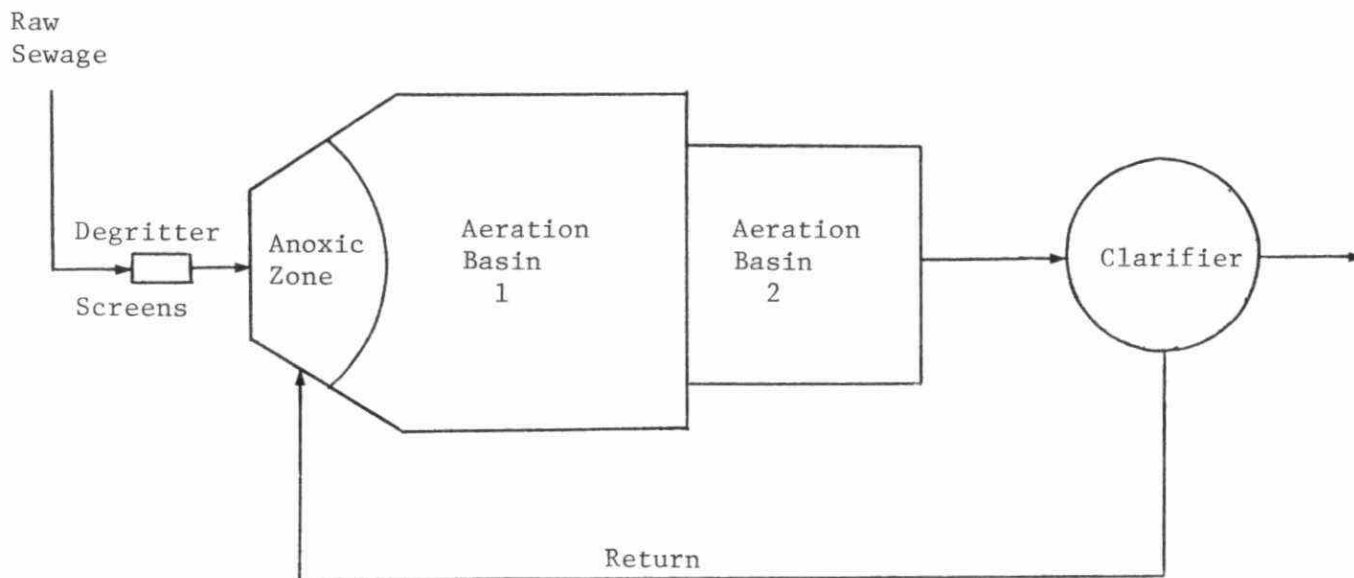


FIGURE 1 (CONT'D)

During the winter months, a bulking mixed liquor developed in the aeration basins and resulted in a high sludge volume index (SVI). This caused secondary clarifier bulking and necessitated lowering the sewage flow to  $136 \text{ m}^3/\text{d}$  ( $30\,000 \text{ Igpd}$ ). The high SVI was diagnosed to be a result of overgrowths of filamentous bacteria caused by very low process food-to-microorganisms ratios (F:M's lower than  $0.1 \text{ kg BOD/kg VSS/d}$ ). Secondary clarifier overflow rates were raised from  $14.7$  to above  $24.5 \text{ m}^3/\text{m}^2$  ( $300$  to  $> 500 \text{ Igpd/ft}^3$ ) by increasing sewage flow on the onset of warmer wastewater temperatures and decreased SVI in summer.

Methanol as a carbon source for denitrification was added to the process only during the summer months. A drop in nitrogen removal ensued and laboratory batch studies confirmed that methanol had inhibited the nitrification reactions. Overall nitrogen reduction across the biological process was 75-85% without the application of methanol. Nitrification reactions proceeded well when mixed liquor DO concentrations were above  $2 \text{ mg/L}$ . Denitrification occurred when the air was turned off and ML DO approached zero.

#### 1.1.3 Barnards's studies

At Pretoria, South Africa, Barnard [5] studied a  $100 \text{ m}^3/\text{d}$  ( $22\,000 \text{ Igpd}$ ) plant for nitrification-denitrification called the "Bardenpho" process. A series of four basins followed by clarification were used for nitrogen removal. The first and third basins were operated in the anoxic state with the second and fourth aerated. Sludge return was provided from both the secondary clarifier and the second basin effluent back to the influent (Figure 1c).

Nitrogen removals exceeding 90% were attained during an 18-month study, but effluent nitrates rose above  $5 \text{ mg/L}$  during winter when basin temperatures dropped below  $20^\circ\text{C}$ . By increasing the basin hydraulic detention time by one hour, increased denitrification was obtained at liquid temperatures approaching  $15^\circ\text{C}$ .

Biological phosphorus removal was observed across the process when denitrification reactions were at their maximum efficiency. A luxury uptake occurred in the fourth basin under aeration when the reactions in the preceding denitrification basin produced considerable orthophosphate release.

#### 1.1.4 Studies by Nicholls

Studies on a modified extended aeration facility for nitrogen removal by Nicholls [6] produced excellent results on domestic sewage without the use of an external carbon source for denitrification (Figure 1d). Two plants of 27 272 m<sup>3</sup>/day (6 MIgpd) and 81 816 m<sup>3</sup>/d (18 MIgpd) capacities were studied under a temperature range of 13-15°C.

The technique of combined nitrification-denitrification was accomplished by turning off four mechanical aerators at the aeration inlet. This created an anoxic zone but maintained enough mixing to keep the mixed liquor suspended solids in motion. This zone functioned as a denitrification area and received raw sewage as well as recirculated secondary clarifier sludge. A large extended aeration process followed this stage.

Suspended solids concentrations in the whole basin varied between 2500 and 6000 mg/L and there were no intervening walls between the various areas. It was demonstrated that as the suspended solids concentration in the basin diminished to 2500 mg/L, denitrification reactions were severely curtailed. Nitrate removals of 98% and total nitrogen reductions of 85% were accomplished with solids retention times (SRT) exceeding 30 days. An aeration basin hydraulic detention time of 20 hours was normal.

#### 1.1.5 Matsche's studies

Studies at the Vienna, Austria, Blumental Waste Treatment Plant by Matsche [7], entailed the use of a 87 273 m<sup>3</sup>/d (19.2 MIgpd) oxidation ditch for nitrogen removal. Nitrification-denitrification was achieved simultaneously by reducing the number of operating mammoth rotors (aerators) along the length of the two aeration channels. By manipulating the number and combination of rotors in operation, aerobic and anoxic pockets were created in various channel locations (Figure 2a).

For this evaluation, two aeration channels were connected in series and three rotors were usually operated in each channel. On occasion, a fourth rotor was used in the first aeration channel to maintain nitrification at peak process loading periods. With the series flow configuration, the mixed liquor DO levels tended to drop to values below 1 mg/L between the rotors in the first aeration channel. With



Figure 2a

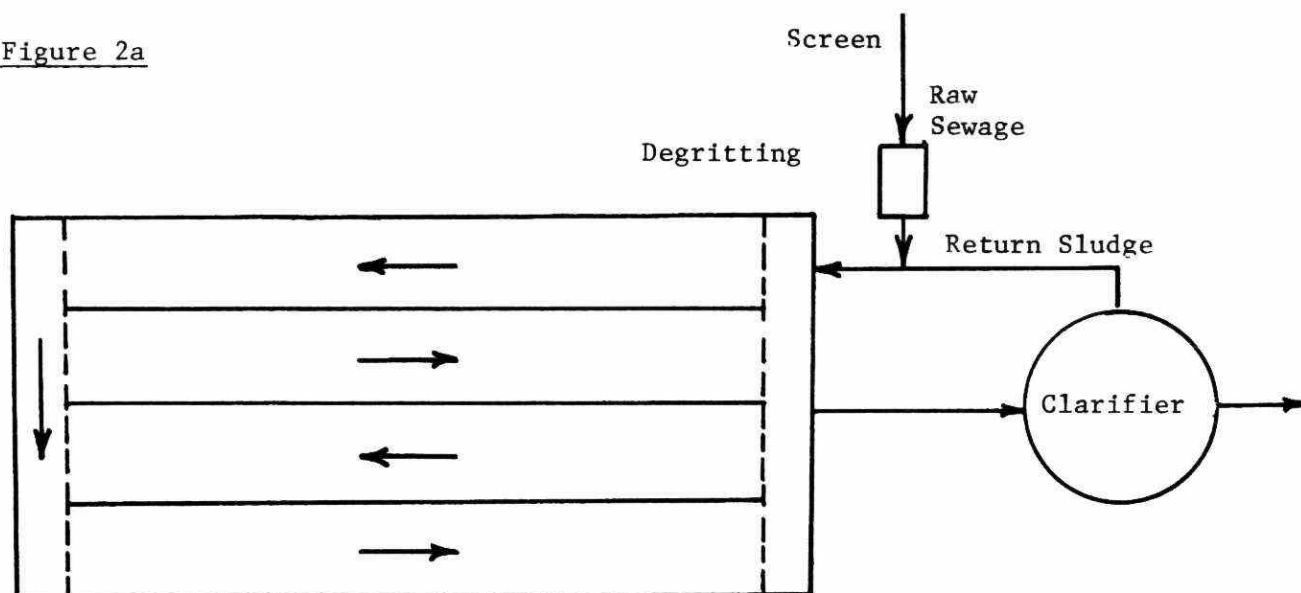


Figure 2b

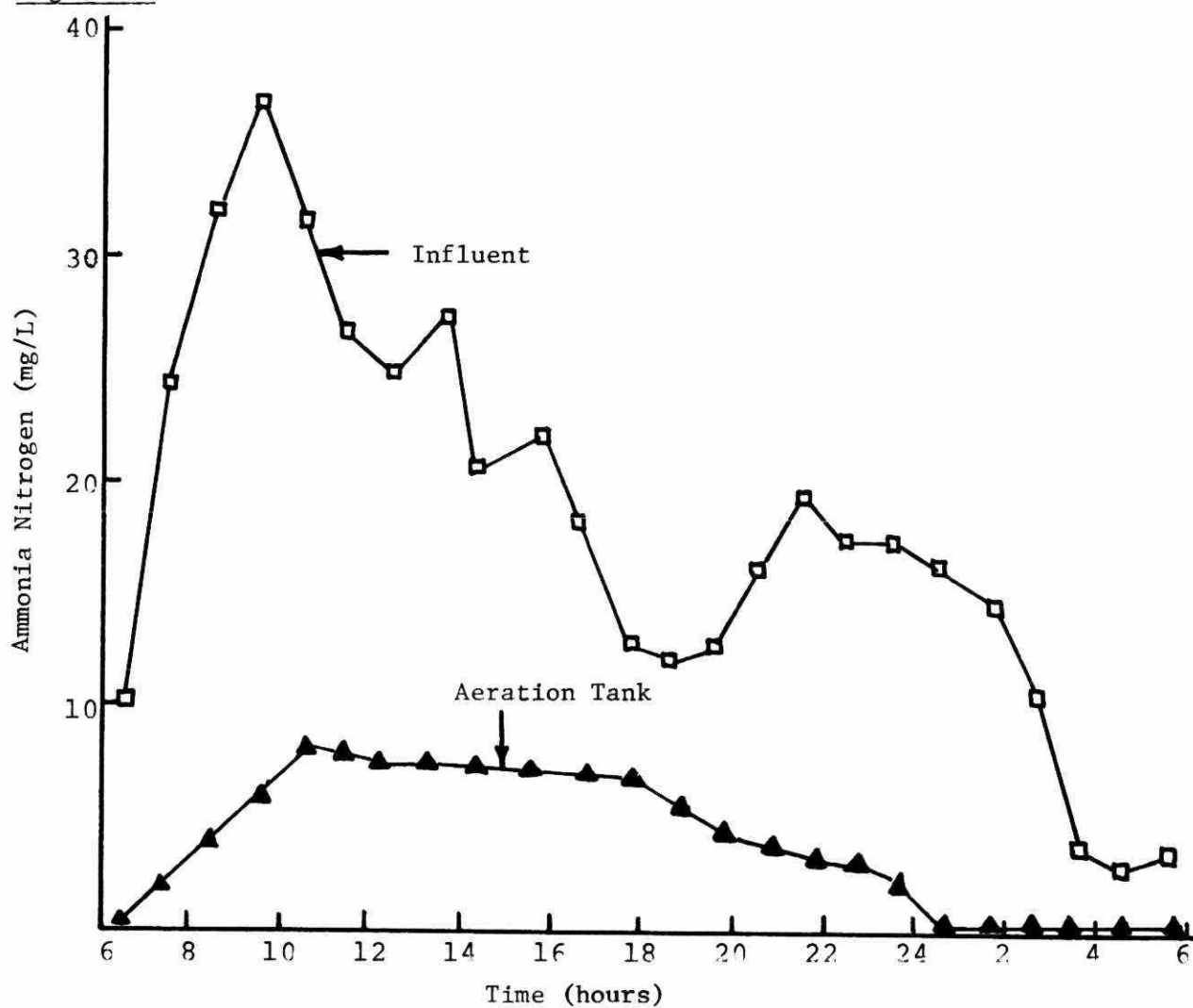


FIGURE 2. THE EFFECTS OF NITROGEN LOADING ON AN OXIDATION DITCH [7]

reduced sewage loading, DO changes were less abrupt and a profile sampling technique showed that wider areas of lower DO depletion occurred. Oxygen utilization rates of the mixed liquors were normally between 80 and 100 mg  $O_2$ /L/h in the first aeration channel and fell, under plug flow conditions and decreased organic loading, to 60 mg/L/h or lower in the second aeration channel.

Depending on the number of rotors in use, variations in aeration ammonia nitrogen concentrations usually followed diurnal variations in process nitrogen loading. As indicated in Figure 2b, aeration ammonia concentrations rose to a value of 8 mg/L by 10 a.m., then decreased gradually until midnight when levels approached zero. The variations were a function of influent ammonia concentrations, which were high at mid-morning and then tapered off throughout the day to values below 10 mg/L. Total nitrogen removals of up to 88% were realized with this process arrangement without a carbon supplement for denitrification. Mixed liquor suspended solids ranged between 3000-7000 mg/L over a period of eight months at liquid temperatures of 12-20°C. The mean monthly organic loading ratio ranged from 0.11 to 0.24 kg BOD/kg aeration solids/day. The SRT's ranged between eight and ten days, and the aeration hydraulic detention time varied from 5.5 to 8.4 hours based on mean monthly data.

## 1.2 Objectives of Present Study

A series of pilot and full-scale studies were conducted by the author at Newmarket, Brampton and Kleinburg, Ontario from October, 1972 to June, 1974 [8]. From these projects it was concluded that the split-return configuration was a viable method of maintaining solids balance in both nitrification and denitrification basins. The value of a post-aeration section between denitrification and secondary clarification stages was also demonstrated.

From these studies it was recommended that additional full-scale evaluations using the split-return arrangement be undertaken. At the same time, the Brampton-Chinguacousy WPCP had been officially acquired from the South Peel Region for research and training. This site, renamed the Ontario Experimental Facility (OEF), was chosen for the study.

The objectives of the present study were:

- 1) to determine ease of operation of a full-scale, single-sludge, split-return arrangement for nitrogen removal;
- 2) to monitor possible effects of temperature upon this process arrangement by making use of seasonal variations;
- 3) to establish the methanol requirements of the denitrification basin which maximize nitrate-nitrogen removal;
- 4) to determine the hydraulic and solids detention times which optimized ammonia oxidation and nitrate removal;
- 5) to observe the effects of ferric chloride application to the system.

### 2.1 The Test Facility

A section of the OEF plant capable of handling up to 18 000 m<sup>3</sup>/d of sewage (4 M gal per day) was chosen for the study. Three main operations were carried out in this section: primary clarification, mechanical aeration, final clarification. A flow diagram of the plant facility is given in Figure 3.

Raw sewage was drawn from a trunk sewer which receives waste from the Town of Brampton and surrounding areas. This waste can be classed as domestic-industrial, with the industrial portion being derived from a variety of small to medium-sized companies. The sewage flows at the intake were variable; the flow into the test facility was maintained within the design limits for process operation by opening and closing a channel gate.

Raw sewage received physical treatment by means of a barminutor and grit removal system, and by primary clarification; the raw sludge was returned to the trunk sewer. Biological treatment was performed in a mechanical aeration chamber having seven bays and was followed by final clarification; the sludge from the clarifiers was returned to the aeration chamber via a metered sludge return system. The final effluent and waste activated sludge were returned to the trunk sewer to be treated, downstream, by the Lakeview WPCP in South Peel. Sewage flow was monitored by a shaped capacitance probe and associated amplifiers connected to a chart recorder. This probe was inserted in the stilling well of a 6-inch Parshall flume situated in the channel between the grit removal facility and the primary clarifiers.

Operation of one leg of the treatment plant was commenced in November, 1973, to test the reliability of the plant equipment for the planned studies and to establish a nitrifying sludge. The first three bays of the remaining leg were separated from bays 4, 5 and 6 by extending divider walls and the seventh bay was isolated in the same way (see Figure 4). The aeration basin comprised bays 1 to 3, each of which had a ten-horsepower mechanical aerator. Three 30-rpm mixers were placed in bays 4 to 6 to facilitate slow mix denitrification, and the seventh bay, which served as a post-aeration chamber prior to final clarification, had a ten-horsepower mechanical aeration unit.

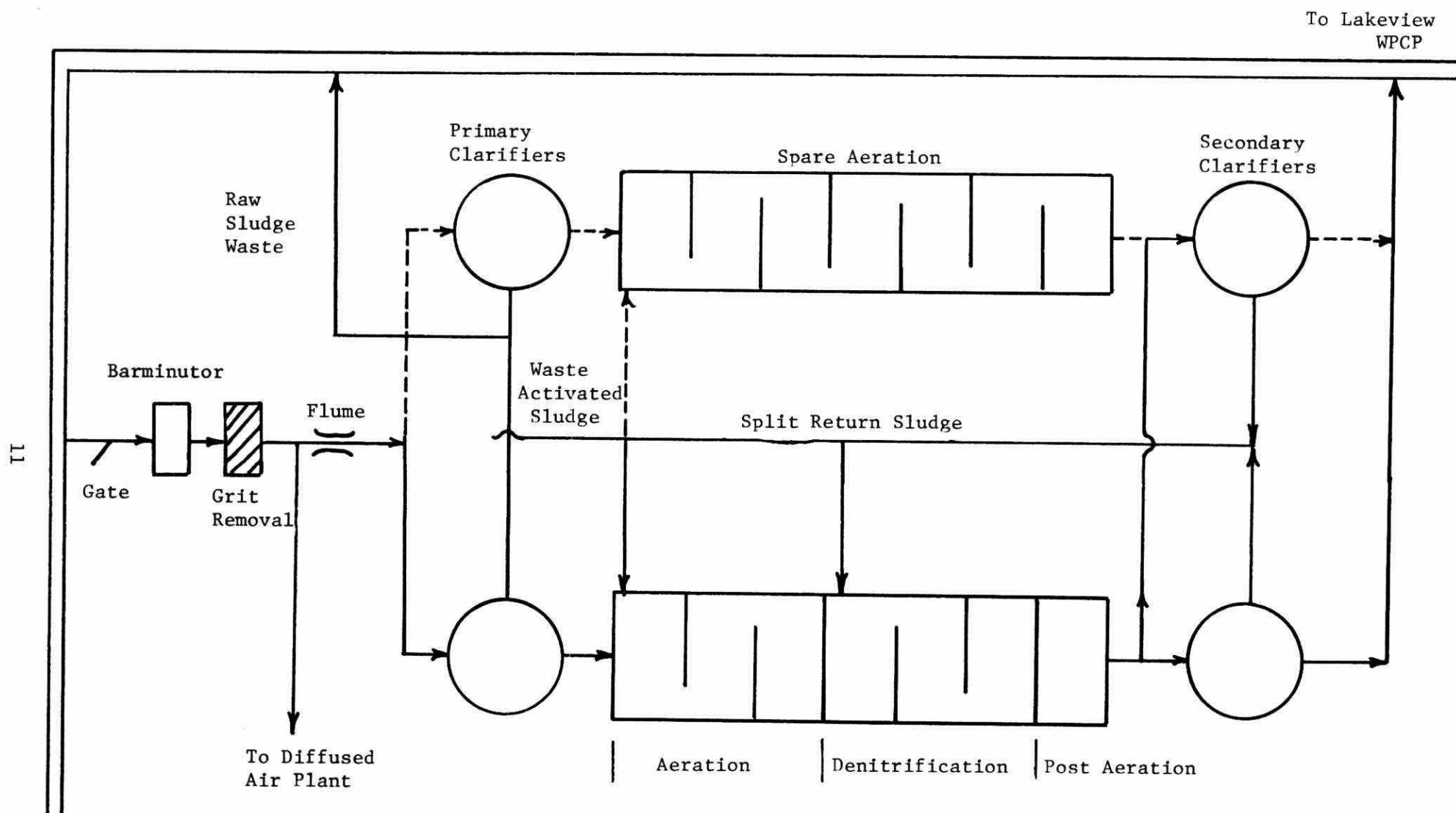


FIGURE 3. OEF NITRIFICATION-DENITRIFICATION FLOW DIAGRAM



PHOTO #1



PHOTO #2

FIGURE 4. OEF NITRIFICATION-DENITRIFICATION CHANNEL

The physical arrangement just described was used for the first four programs of this study. Alterations were made to change basin volumes in subsequent programs. Further details of plant basin volumes are given in Table 1.

TABLE 1. PLANT BASIN VOLUMES

---

<u>Volumes of Major Elements</u>	
Primary Clarifiers	- 568 m <sup>3</sup> (125 000 IG)
Final Clarifiers	- 795 m <sup>3</sup> (175 000 IG)
Total Nitrogen	
Removal Basin	- 2273 m <sup>3</sup> (500 000 IG)
<u>Volumes in Study Programs</u>	
<u>Programs 1-4</u>	
Nitrification (Aeration)	- 973 m <sup>3</sup> (214 000 IG)
Denitrification	- 973 m <sup>3</sup> (214 000 IG)
Post-Aeration	- 327 m <sup>3</sup> (72 000 IG)
<u>Programs 5 (a), (b) and (c)</u>	
Nitrification	- 1300 m <sup>3</sup> (286 000 IG)
Denitrification	- 327 m <sup>3</sup> (72 000 IG)
Post-Aeration	- 655 m <sup>3</sup> (144 000 IG)
<u>Program 6</u>	
Nitrification	- 1300 m <sup>3</sup> (286 000 IG)
Denitrification	- 655 m <sup>3</sup> (144 000 IG)
Post-Aeration	- 327 m <sup>3</sup> (72 000 IG)

---

## 2.2 Study Design

The OEF project was classified into a number of programs, each of which entailed a minimum of 25 days process operation. These programs were as follows:

- Program 1  
(July 22-Oct. 4/74)
- Basin hydraulic detention
    - aeration 8 h
    - denitrification 8 h
    - post-aeration 3 h
- Mean liquid temperature of 18°C.
- Program 2  
(Oct. 7-Nov. 29/74)
- Basin hydraulic detention
    - aeration 6 h
    - denitrification 6 h
    - post-aeration 2 h
- Mean Liquid temperature of 16°C.
- Program 3  
(Jan. - Mar. 15/75)
- Same as Program 2 but with a mean liquid temperature of 10°C.
- Program 4
- Same as Program 1 but with a mean liquid temperature of 10°C.
- Peroxide Study  
(May/75)
- Same as Program 4 but with the addition of hydrogen peroxide to the process in an attempt to control bulking sludge.
- Program 5(a)  
(June 17-July 25/75)
- Basin hydraulic detention
    - aeration 8 h
    - denitrification 2 h
    - post-aeration 4 h
    - methanol/nitrate ratio 3:1 (w/w)
- Mean liquid temperature of 20°C.
- Program 5(b)  
(July 28-Aug. 29/75)
- Basin hydraulic detention
    - aeration 10 h
    - denitrification 2.5 h
    - post-aeration 5 h
    - methanol/nitrate ratio 6:1 (w/w)
- Mean liquid temperature of 22°C.
- Program 5(c)  
(Sept. 2-26/75)
- Basin hydraulic detention
    - aeration 10.5 h
    - denitrification 2.6 h
    - post-aeration 5.2 h
- No methanol was added to denitrification during this sequence.
- Mean liquid temperature of 18°C.
- Program 6  
(Sept. 29-Dec. 18/75)
- Basin hydraulic detention
    - aeration 10 h
    - denitrification 4 h
    - post-aeration 2 h
- Ferric chloride was added to the post-aeration basin.



The overall process parameters applied in Programs 1 to 6 were as follows:

- a) aeration F:M ratio maintained at 0.15 to 0.3 kg BOD/kg VSS/d;
- b) aeration ML dissolved oxygen between 2-4 mg/L;
- c) secondary clarifier sludge return to basins was 100 to 200% of plant flow and split between aeration and denitrification processes;
- d) mixed liquor suspended solids concentration in the denitrification basin maintained between 4500-6000 mg/L;
- e) except for Programs 5(a)-(c), methanol was added in amounts sufficient to maintain 1 mg/L or less of nitrate nitrogen in the denitrification basin.

## 2.3 Routine Sampling and Analysis

### 2.3.1 Sampling

Two sets of samples were taken on a routine basis. The first set consisted of 24-hour composite samples of raw sewage, and of primary and secondary clarifier effluents, taken five days per week. Portions of these composites were filtered for analysis of their soluble components. The second set were filtered composite grab samples of the mixed liquor from the aeration and denitrification basins. These were taken on a daily basis and analysed for the dissolved chemical species specified in Section 2.3.2.

### 2.3.2 Analyses

A number of on-site tests, needed for process control purposes, were performed daily at the OEF laboratory. These were:

- mixed liquor  $\frac{1}{2}$  h settling tests, suspended and volatile solids, dissolved oxygen and utilization rates, pH;
- analyses of mixed liquor filtrates for ammonia and nitrate nitrogen.

A more comprehensive series of chemical analyses were carried out daily on samples taken to the Ministry of Environment (MOE) Central Laboratory. These analyses were:

- i) Unfiltered samples - biochemical oxygen demand (BOD<sub>5</sub>), chemical oxygen demand (COD), total phosphorus, alkalinity, total Kjeldahl nitrogen (TKN) and suspended solids.
- ii) Filtered samples - inorganic and organic carbon, phosphorus, ammonia, nitrite and nitrate nitrogen.

Secondary effluent total oxygen demand (TOD) was determined by a formula currently used to calculate TOD discharges for nitrification plants in southwestern Ontario [9]:

$$\text{TOD} = \text{NOD} + \text{CARBOD}$$

where NOD is the oxygen demand due to nitrogen containing species, and CARBOD is the oxygen demand of the organic carbon species. NOD is related to the parameter total Kjeldahl nitrogen (TKN):

$$\text{NOD} = 4.57 \times \text{TKN}$$

and CARBOD is defined as the 20-day biochemical oxygen demand (BOD<sub>20</sub>). The BOD value commonly measured in sewage treatment is, however, the five-day determination (BOD<sub>5</sub>). A conversion factor R, the ratio of BOD<sub>20</sub>/ BOD<sub>5</sub> for the same sample of sewage (an experimentally determined value) is used to convert BOD<sub>5</sub> to CARBOD:

$$\therefore \text{TOD} = 4.57 \times \text{TKN} + R \times \text{BOD}_5$$

#### 2.4 Facility Operations and Non-Scheduled Evaluations

The nitrification-denitrification process was initiated in June, 1974, when the established aeration mixed liquor (ML) was transferred from the spare aeration section to the test nitrification-denitrification basin. During the first three weeks of observation considerable backflow of denitrification mixed liquor to the aeration basin was observed. As a consequence aeration ML concentration rose above the desired level without the need for recycle flow, and low DO concentration was observed.

The mixed liquor was pumped to the spare aeration section and the divider walls were sealed with the exception of a 0.6 x 2.9 m (2' x 8') opening located 1.8 m (6') below the liquid surface. The nitrogen removal process was then restarted and evaluated until early September. During this period an improvement occurred in aeration nitrification, but solids maintenance was difficult as some denitrification mixed liquor still appeared to be backing up into the aeration basin.

The process mixed liquor was again transferred to the spare aeration basin, only this time the wall divisions between the aeration, denitrification, and post-aeration basins were completely closed except for a one-way flapgate. Following this modification the nitrification-denitrification system operated well without backmixing.

A volume of 1/6th to 1/8th of the total return sludge was then required to maintain aeration solids between 1500 and 2500 mg/L. The remainder of the return sludge flow was directed to the denitrification basin, and wasting of return sludge was usually implemented when denitrification solids rose above 5500 mg/L.

Once the principal alterations were made to the aeration and denitrification basins, attention was directed to influent flow control. Hourly sampling revealed a rise in aeration ammonia concentration caused by periods of simultaneous high flow and influent TKN levels. In an effort to reduce these peaks, the operational staff experimented with influent gate control. They found that the required daily average could be attained with a flow pattern that resembled a typical average dry weather flow (ADWF), by setting the late morning flow to 40% above the daily mean.

Most of these aforementioned adjustments were made prior to, and during, Program 1. As a result, this study period had to be extended in order to gain enough data obtained under stable conditions.

The project was interrupted following Program 4 to treat a bulking sludge condition which necessitated the use of an additional secondary clarifier to avoid loss of solids. These trials were conducted during the month of May, 1975.

Intermittent applications of hydrogen peroxide ( $H_2O_2$ ), at concentrations ranging from 25 to 127 mg/L, were applied at the post-

aeration stage and (eventually) to the return sludge. A loss of nitrification efficiency resulted with  $H_2O_2$  application; consequently, it was discontinued.

Program 5 involved expansion of the aeration section to four bays, and the post-aeration to two bays, leaving one denitrification bay. Step feeding of primary effluent to the first three bays of the aeration basin was also implemented to equalize loading. Nitrogen removal under these revised conditions was evaluated with and without methanol application.

During Program 5(c), a pilot-scale 454 L (100 gal)/minute screen centrifuge concentrator was tested for its effect on the settling of denitrification mixed liquor. The unit was evaluated using, alternately, 160 and 325-mesh screens. Sludge settling rate determinations and microscopic observations were made routinely throughout this evaluation.

In Program 6, the denitrification section was enlarged to two basins and the post-aeration was reduced to one. Ferric chloride was added to the post-aeration, initially at 8.0 mg/L and then gradually increased to 30 mg/L (as Fe) over two months, in an effort to provide phosphorus removal and possibly a lower sludge volume index (SVI).

Due to the bulking nature of the system mixed liquor, sedimentation tests were performed on the secondary clarifier. The reference point of blanket-loss, which was defined as the condition when the clarifier sludge-blanket rose to within 60 cm (2 ft) of the surface, was used as a guide. Evaluations using the point of blanket-loss were conducted during and following the nitrogen removal study.

### 3 RESULTS AND DISCUSSION

#### 3.1 Primary Clarification Stage

As can be seen from Table 2, higher-than-normal BOD and SS reductions occurred through primary clarification. These results may have been due to the relatively low clarifier surface loading rates used in this study; between 16 and 21 m<sup>3</sup>/m<sup>2</sup>/d (330 and 420 Igpd/ft<sup>2</sup>)

The mean percentage reductions in sewage quality parameters (BOD<sub>5</sub>, SS, TKN) achieved in primary clarification for each program are shown in Table 3. Mean suspended solids (SS) reductions ranged from 69 to 84%; removal of nitrogen, which ranged from 10 to 41% of total Kjeldahl nitrogen (TKN), paralleled the removal of SS. The reduction in biochemical oxygen demand (BOD<sub>5</sub>) ranged from 46 to 66%, with the removal efficiency being primarily a function of the SS concentration in the clarifier influent/effluent. These values may be compared to 30% and 43% BOD<sub>5</sub> and SS removal, respectively, obtained in conventional MOE activated sludge plants operated at a loading rate of 34 m<sup>3</sup>/m<sup>2</sup>/d (700 Igpd/ft<sup>2</sup>) [10].

#### 3.2 Nitrification

##### 3.2.1 Effects of DO concentrations

As stated in Section 2.4, nitrification was lacking in the aeration basin during the startup period due to backmixing of denitrification mixed liquor. At peak loading periods aeration DO concentrations ranged from less than 1 mg/L in the first bay to no more than 2 mg/L in the third bay. Under these conditions, a mean of 55% nitrification was produced prior to the denitrification basin.

Following the first alteration to the basin divider walls, a noticeable rise in mean aeration DO, to between 2 and 5 mg/L at diurnal peak loading, was observed. An immediate improvement in aeration nitrification efficiency to more than 80% resulted.

Dissolved oxygen surveys were conducted twice daily, morning and afternoon, during the startup period and the early weeks of Program 1, and revealed a relationship between aeration DO and nitrification rate. These results are presented in Figure 5. Based on the weight of available aeration MLSS, a significant rise in ammonia oxidation rate occurred upon

TABLE 2. OEF PRIMARY STAGE ANALYTICAL DATA (in mg/L except where indicated)

Program	Raw Sewage						Primary Effluent					Filtrate					
	Flow (m <sup>3</sup> /d)	BOD	COD	Tot. P	TKN	SS	BOD	SS	Tot. P	TKN	COD	Sol. P	Alk.	NH <sub>3</sub>	NO <sub>2</sub> <sup>-</sup>	NO <sub>3</sub> <sup>-</sup>	TC
1	3123	166	643	7.6	38	223	89	70	6.3	28	249	3.9	215	16	.03	.2	99
2	3751	206	649	9.2	40	346	99	89	7.1	30	250	4.9	230	21	.03	.2	92
3	3737	162	579	7.1	32	239	85	72	6.2	29	238	3.5	227	18	.06	.09	99
4	3146	161	483	6.9	29	256	73	53	5.2	21	188	3.3	235	13	.06	.10	96
5(a)	3787	242	694	9.1	36	424	86	66	5.2	24	192	3.7	236	19	.02	.1	91
5(b)	3191	234	727	9.0	38	464	79	78	5.2	22	207	3.4	204	22	.03	.05	93
5(c)	2973	273	694	8.7	39	493	120	89	5.3	27	241	4.1	221	18	.03	.05	105
6	3214	256	614	9.0	43	481	101	89	5.4	27	225	4.0	247	19	.02	.26	111

TABLE 3. EFFECTS OF VARIOUS OPERATING PARAMETERS ON PRIMARY CLARIFICATION  
(all quantities are expressed as mean values for each program)

Program	Clarifier Conditions				Sewage Quality Parameters		
	Sewage Temperature °C	Solids Loading kg SS/m <sup>2</sup> /d (lbs SS/ft <sup>2</sup> /d)	Detention h	Surface Loading m <sup>3</sup> /m <sup>2</sup> (lgpd/ft <sup>2</sup> )	BOD	SS % Removal	TKN
1	18	3.81 (0.78)	4.4	17.1 (350)	46	69	26
2	16	7.52 (1.54)	3.6	20.6 (420)	52	74	25
3	10	4.88 (1.00)	3.6	20.5 (418)	48	70	10
4	10	4.39 (0.90)	4.3	17.2 (352)	55	79	28
5(a)	20	8.74 (1.79)	3.6	20.7 (424)	64	84	35
5(b)	22	8.10 (1.66)	4.3	17.5 (358)	66	83	42
5(c)	18	8.01 (1.64)	4.6	16.3 (333)	56	82	32
6	16	8.44 (1.73)	4.9	17.6 (360)	56	81	37
Mean of Means	16	6.69 (1.37)	4.2	18.4 (377)	55	78	29

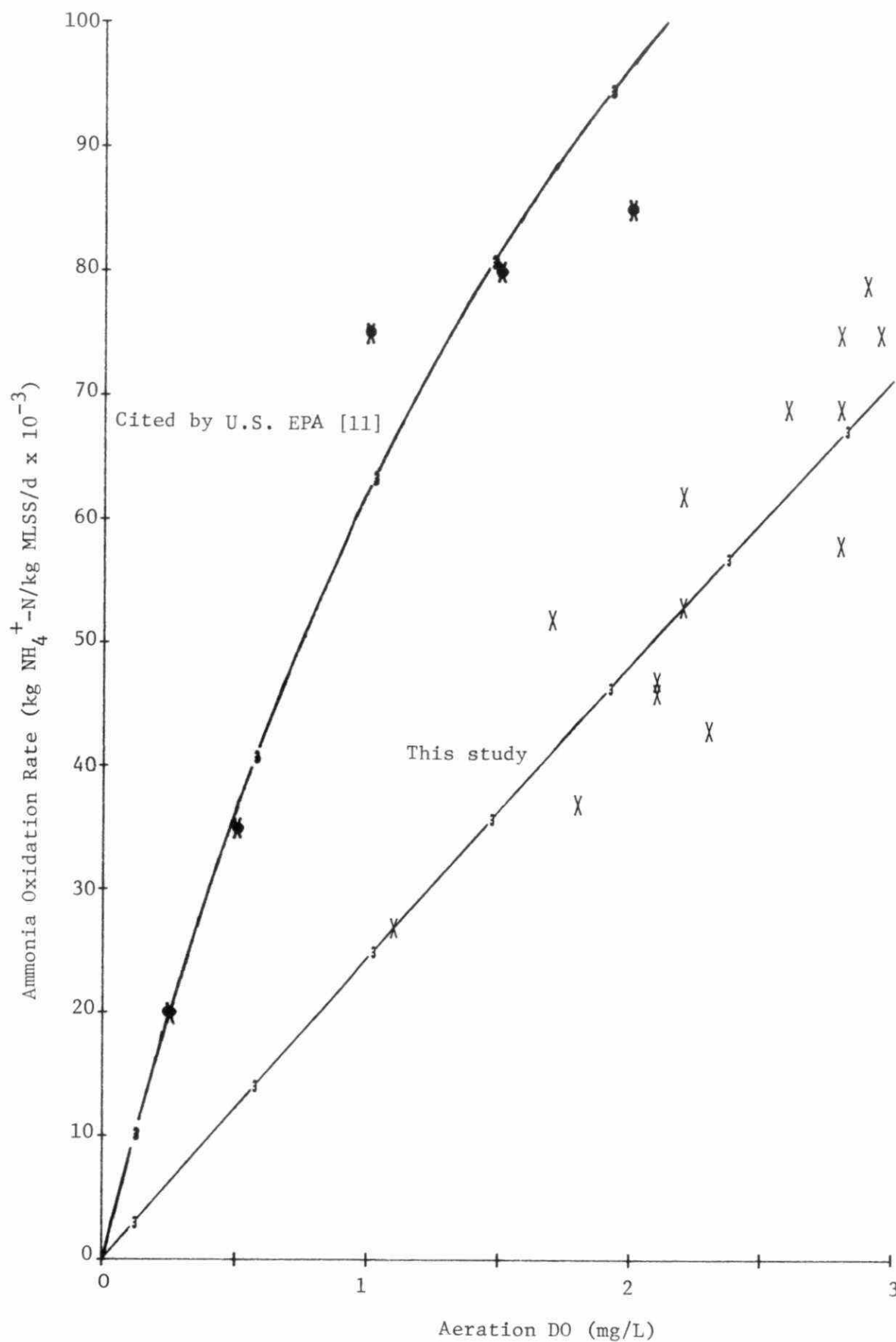


FIGURE 5. INFLUENCE OF DISSOLVED OXYGEN ON AMMONIA OXIDATION RATE



raising mean aeration DO from 1 to 3 mg/L. A nitrification rate of approximately 0.023 kg  $\text{NH}_4\text{-N}$  oxidation/kg MLSS/d was produced at an aeration DO of 1 mg/L. An aeration DO of 3 mg/L produced a nitrification rate of 0.07.

Data from studies by Nagel and Haworth, equating aeration DO with nitrification rate, are presented in the U.S. Environmental Protection Agency's (EPA) nitrogen design manual [11]. The OEF study data are plotted and compared with the results in the U.S. study in Figure 5. The OEF results were considerably lower at all temperature ranges than those in the U.S. EPA manual [11].

On maintaining aeration DO levels above 3 mg/L in Programs 2 to 4, some improved nitrification occurred. As shown in Table 4, aeration filtrate ammonia nitrogen levels dropped from a mean of 3.8 mg/L in Program 1, to 1.0 mg/L in Program 4. Step feeding of the primary effluent to the first three bays in Programs 5 and 6 produced uniform DO throughout the aeration basin at much higher liquid temperatures than either Programs 3 or 4 (Table 5).

### 3.2.2 Loading effects

As discussed previously (Section 2.4), aeration ammonia levels varied with fluctuations in process loading. Increases in flow and influent ammonia concentration in the late morning were followed by raised aeration ammonia level (up to 8 mg/L) in the afternoon and evening (see Figure 6). The residual ammonia was oxidized in the post-aeration basin; consequently, values below 2 mg/L were observed in secondary effluent. Denitrification effluent ammonia concentrations fell between those found in the aeration and post-aeration basins, probably due to dilution by the return sludge.

Residual ammonia concentrations at the aeration basin were generally higher during Program 1 (see Table 4). Improved aeration nitrification occurred in the subsequent programs with better control over flow peaking. Further evidence of this is shown in the final column of Table 5 where aeration and total process ammonia removals are compared. Close to complete nitrification was obtained in the aeration basin through Programs 4 to 6, with the exception of 5(c) (see Figure 7).

TABLE 4. ANALYTICAL DATA FROM AERATION AND DENITRIFICATION BASINS  
(Mean values in mg/L)

Program	Aeration							Denitrification						
	M.L.		Filtrate					M.L.		Filtrate				
	SS	VSS	Sol. P	NH <sub>3</sub>	NO <sub>2</sub> <sup>-</sup>	NO <sub>3</sub> <sup>-</sup>	TC	SS	VSS	SOL. P	NH <sub>3</sub>	NO <sub>2</sub> <sup>-</sup>	NO <sub>3</sub> <sup>-</sup>	TC
1	2311	1841	2.9	3.8	1.0	7.9	45	4241	3594	3.4	.8	.25	1.0	48
2	2268	1592	2.3	2.5	.4	7.4	47	4647	3386	4.0	.8	.25	1.7	52
3	2201	1768	2.4	2.3	.3	8.5	49	4723	3784	3.0	.7	.07	.9	50
4	1818	1455	2.6	1.0	.04	10.3	48	5551	4477	3.4	.2	.07	.8	55
5(a)	1446	1370	2.7	0.9	.8	9.9	40	4150	3317	4.1	.6	.12	2.8	51
5(b)	1968	1569	3.4	1.6	.05	7.9	43	5016	3446	4.5	1.3	.02	2.1	49
5(c)	1507	1199	4.0	2.8	.02	11.1	43	4770	3466	3.7	2.3	.07	4.5	47
6	1828	1309	1.1	1.4	.07	8.1	46	4975	3595	.9	1.0	.06	1.5	46

TABLE 5. PROCESS CONDITIONS AND NITRIFICATION EFFICIENCY

Program	Sewage Temp. °C	GSA (days)	SRT (days)		Aeration F/M Ratio BOD <sub>5</sub> /VSS	Hours Aeration Detention		% Ammonia Removal	
			Total	Aeration		Nominal	Actual*	Aeration	Total
1	18	10	63	34	0.16	7.9	6.7	76	95
2	16	6.6	67	13	0.24	6.2	4.5	88	96
3	10	7.9	222	99	0.19	6.2	4.6	87	96
4	10	11	41	16	0.16	7.4	6.5	92	98
5(a)	20	7.5	122	93	0.18	8.2	7.0	95	98
5(b)	22	10	213	166	0.12	9.8	8.1	93	94
5(c)	18	7.4	70	54	0.23	10	9.0	84	87
6 (Fe <sup>+3</sup> )**	16	8.3	16	9	0.18	9.7	8.8	93	95

\*Detention time with recycle flow.

\*\*Ferric chloride added to process.

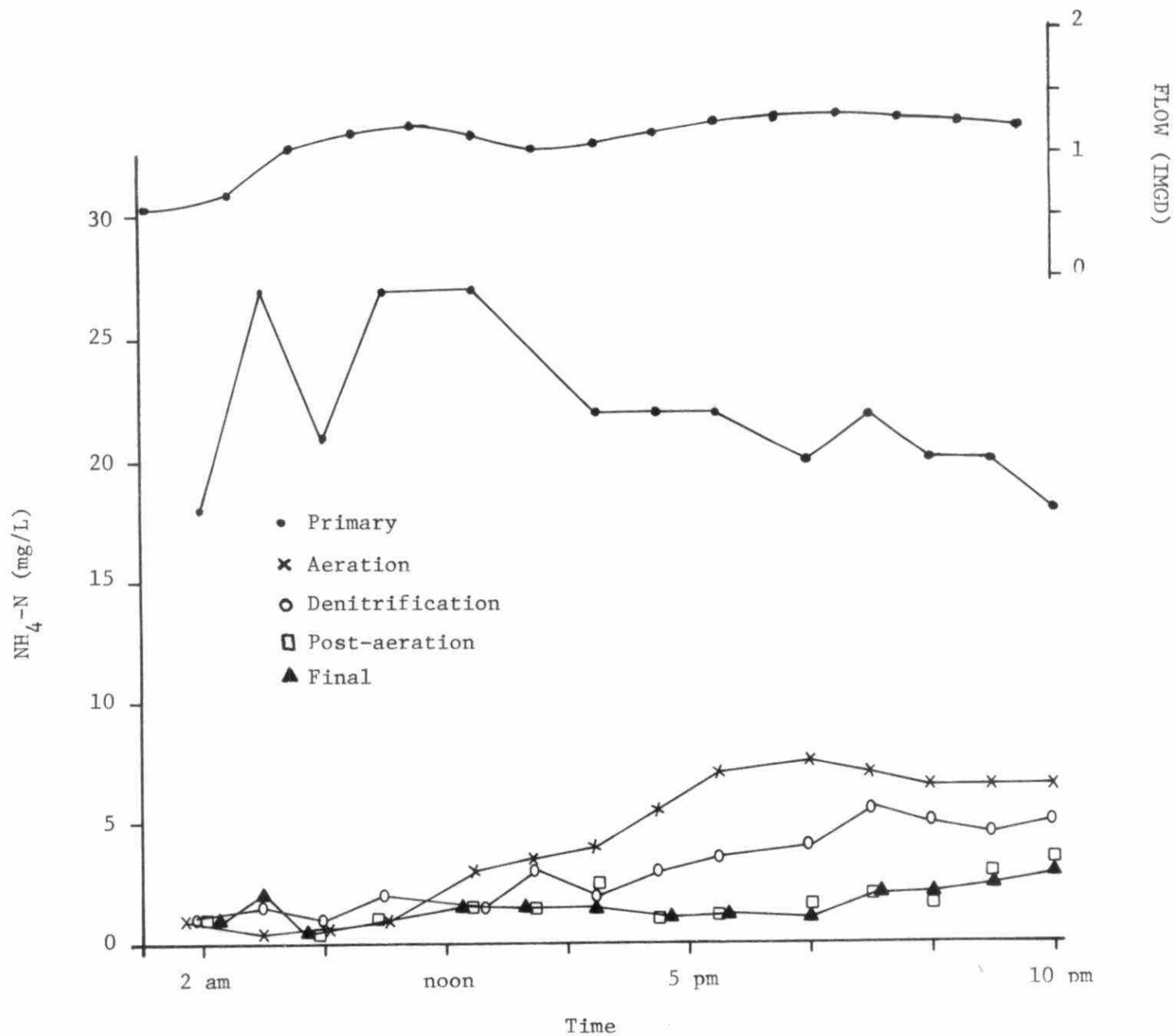


FIGURE 6. EFFECTS OF INFLUENT FLOW AND AMMONIA CONCENTRATION ON OEF NITRIFICATION-DENITRIFICATION SYSTEM

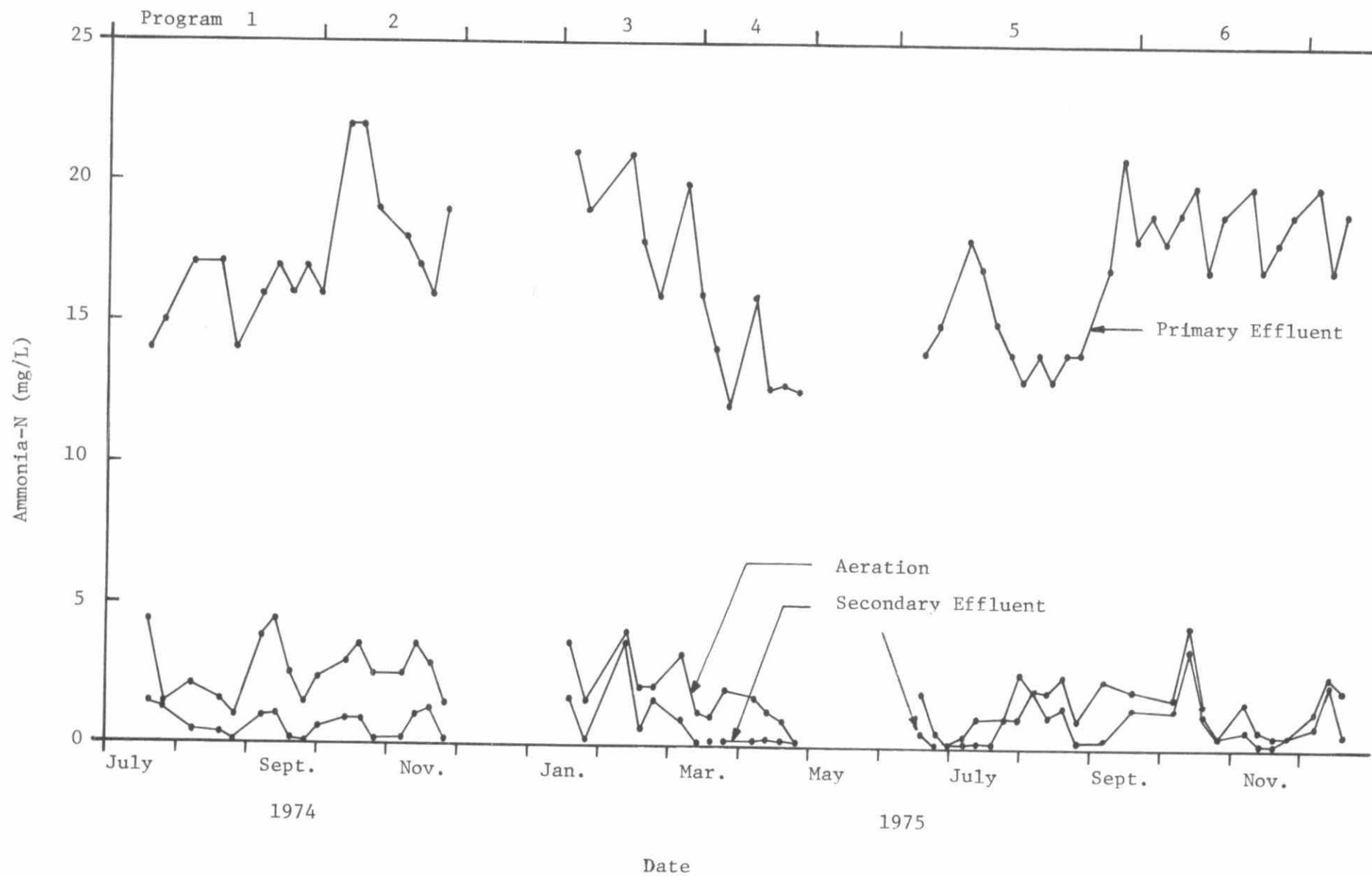


FIGURE 7. AMMONIA-N DISTRIBUTION

### 3.2.3 Effects of sludge age, SRT, and temperature

The definition of sludge age provided by Gould and cited in [12] was used in this study.

$$\text{Gould Sludge Age (GSA)} = \frac{\text{wt. aeration solids}}{\text{wt. primary solids added/day}} \text{ (days)}$$

Solids retention time (SRT) was also determined in an attempt to establish a relationship with nitrification efficiency. SRT was defined as:

$$\text{SRT} = \frac{\text{wt. system solids}}{\text{wt. wasted and effluent solids/day}} \text{ (days)}$$

The system solids in this expression were defined as aeration plus denitrification plus post-aeration when referring to total SRT. A partial or aeration SRT was also tested in which only the aeration solids weight was used.

For the most part, the widest range of nitrification efficiencies and temperature occurred in Programs 1 to 4 (Table 5). Analyses of data for these programs revealed a relationship between GSA and percent removal of ammonia which was independent of temperature (Figure 8). Ammonia removal of 95% resulted at six days compared to 80% at two days GSA.

Similar data computations were tried in an attempt to equate SRT to nitrification efficiency and rate, but no correlations were found with either total or aeration SRT. This finding is illustrated in Table 5 by a comparison of the mean program values of SRT and ammonia removal. It should also be noted that no correlation was found between nitrification efficiency and F/M ratio at mean program values of 0.12 to 0.24 kg BOD<sub>5</sub>/kg aeration VSS.

Unlike the results reported by Bishop [4] and Matsche [7], wastewater temperature did not apparently influence nitrification within the GSA and SRT ranges of the various programs (Table 5). The lowest wastewater temperatures, as low as 5°C on some days, occurred in Program 3. Nitrification rate (wt. ammonia removed/wt. aeration MLVSS/day) generally followed ammonia loading, particularly in the first four programs (Table 6). At the mean program temperatures, ranging from 10 to 22°C, nitrification rates similar to those reported by Wuhrmann [3] and Bishop [4] at 14 and 15°C, respectively, were observed. Lower values (0.006 to 0.012) were calculated from data reported by Barnard [5], Nicholls [6] and Matsche [7].

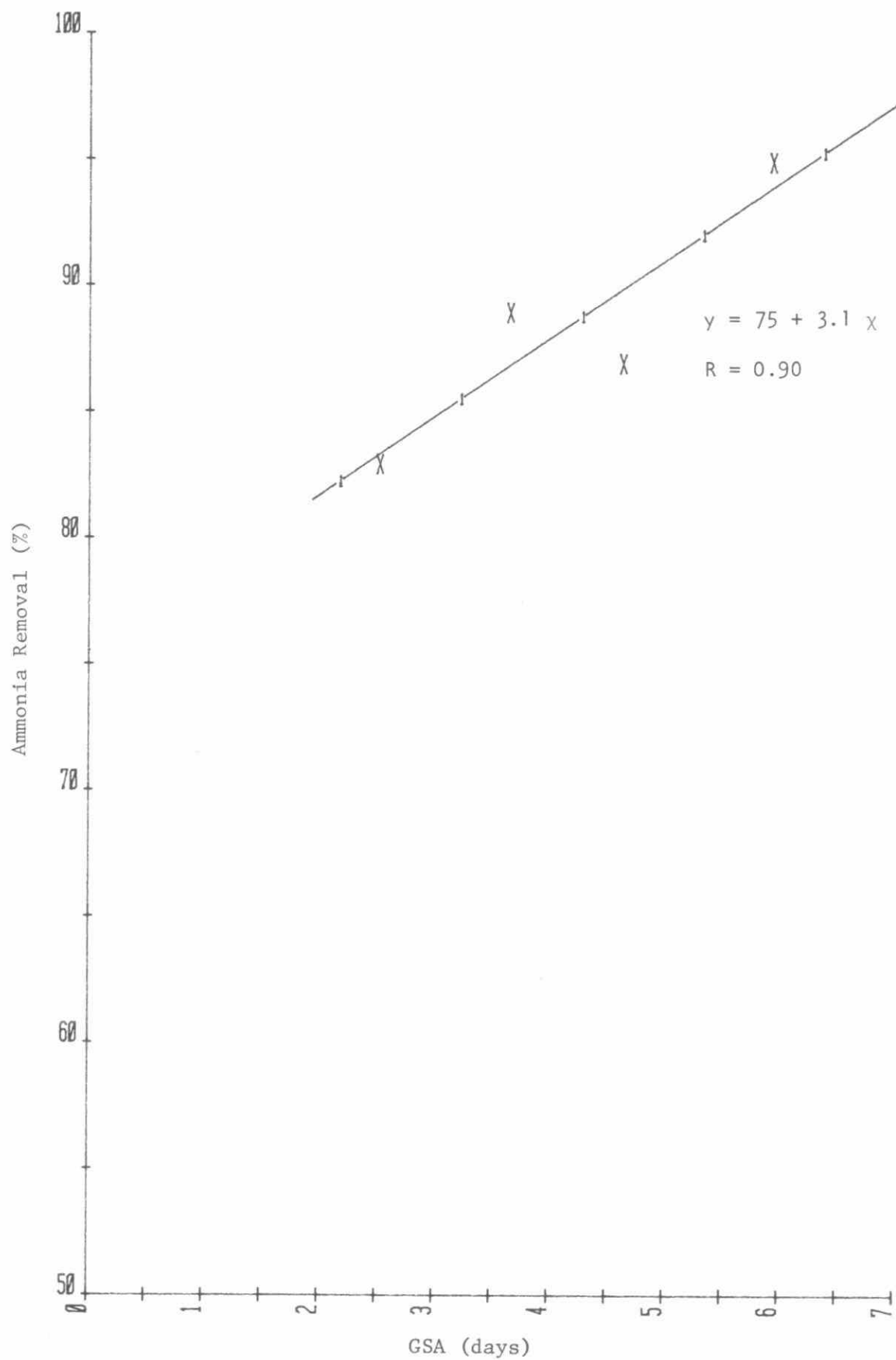


FIGURE 8. RELATIONSHIP BETWEEN AMMONIA REMOVAL AND GSA, PROGRAMS 1 TO 4

TABLE 6. COMPARISON OF NITRIFICATION RATES

Program or Reference	Temp. °C	NH <sub>4</sub> -N Applied kg/day (lbs/day)	Nitrification Rate (kg NH <sub>4</sub> <sup>+</sup> -N removed per kg aeration VSS)	Notes
1	18	49.9 (110)	0.027	aeration section of combined system
2	16	78.5 (173)	0.049	
3	10	67.1 (148)	0.037	
4	10	40.8 ( 90)	0.029	
5(a)	20	71.7 (158)	0.040	
5(b)	22	69.9 (154)	0.032	
5(c)	18	53.5 (118)	0.032	
6	16	60.8 (134)	0.034	
Wuhrmann [3]	14		0.032	nitrification system
Bishop [4]	15		0.030	total basin cycled aerobic/anaerobic
	25		0.089	
	27		0.110	
	26		0.120	
Barnard [5]	14		0.012	calculation based on assumed 80% volatile content of ML
Nicholls [6]	13-25		0.007	same
Matsche [7]	12		0.006	same
	20		0.012	



#### 3.2.4 Consistency of effluent ammonia

Effluent ammonia probability plots for all the programs are shown in Figures 9 and 10. Consistent levels below 0.3 mg/L were produced in Program 4 at the lowest ammonia loading rate of 41 kg/d (90 lbs/day) and lowest mean wastewater temperature (10°C) (Table 6). Based on the total aeration and post-aeration VSS, a mean loading rate of 0.014 kg  $\text{NH}_4^+\text{-N/kg VSS/day}$  was derived in this program, compared to values of 0.017 and 0.03 for Programs 1 and 3, respectively. Effluent ammonia distribution in Programs 1 and 2 indicated similar patterns, with the results showing 80% of values below 1.2 mg/L.

In terms of overall ammonia oxidation consistency, later programs (5(b) to 6) showed no improvement with increased aeration volume (Figure 10). Program 5(a) produced 80% of effluent ammonia-N values below 0.6 mg/L, which compares well with results of Program 4.

A slight rise in effluent ammonia-N was observed in Programs 5(b), 5(c), and 6 as indicated by an increase in mean values to between 1 and 2 mg/L. Furthermore, there was a greater variation in values in the latter programs (Figure 10). Only a slight effect on nitrification was noted when ferric chloride was added in program 6. A mean effluent concentration of 1.2 mg/L ammonia resulted, with 80% of the values below 2.4 mg/L (Figure 10 and Table 7).

#### 3.2.5 Inorganic carbon removal

Inorganic carbon analyses of process effluents and mixed liquor revealed, for the most part, poor correlation between TIC and ammonia oxidation, at high nitrification efficiency. The exception occurred in Program 1, where the widest variation in nitrification was observed. A 10% rise in soluble inorganic carbon utilization was attained as aeration nitrification efficiency rose from 72 to 100% (Figure 11). These results were less dramatic than those obtained previously during pilot-plant studies at Brampton, where a 30% increase in TIC removal resulted through similar increases in nitrification efficiency [8].

#### 3.3 Denitrification

As with nitrification, diurnal flow variation also influenced effluent nitrate concentrations. Nitrate-N levels rose with the influent

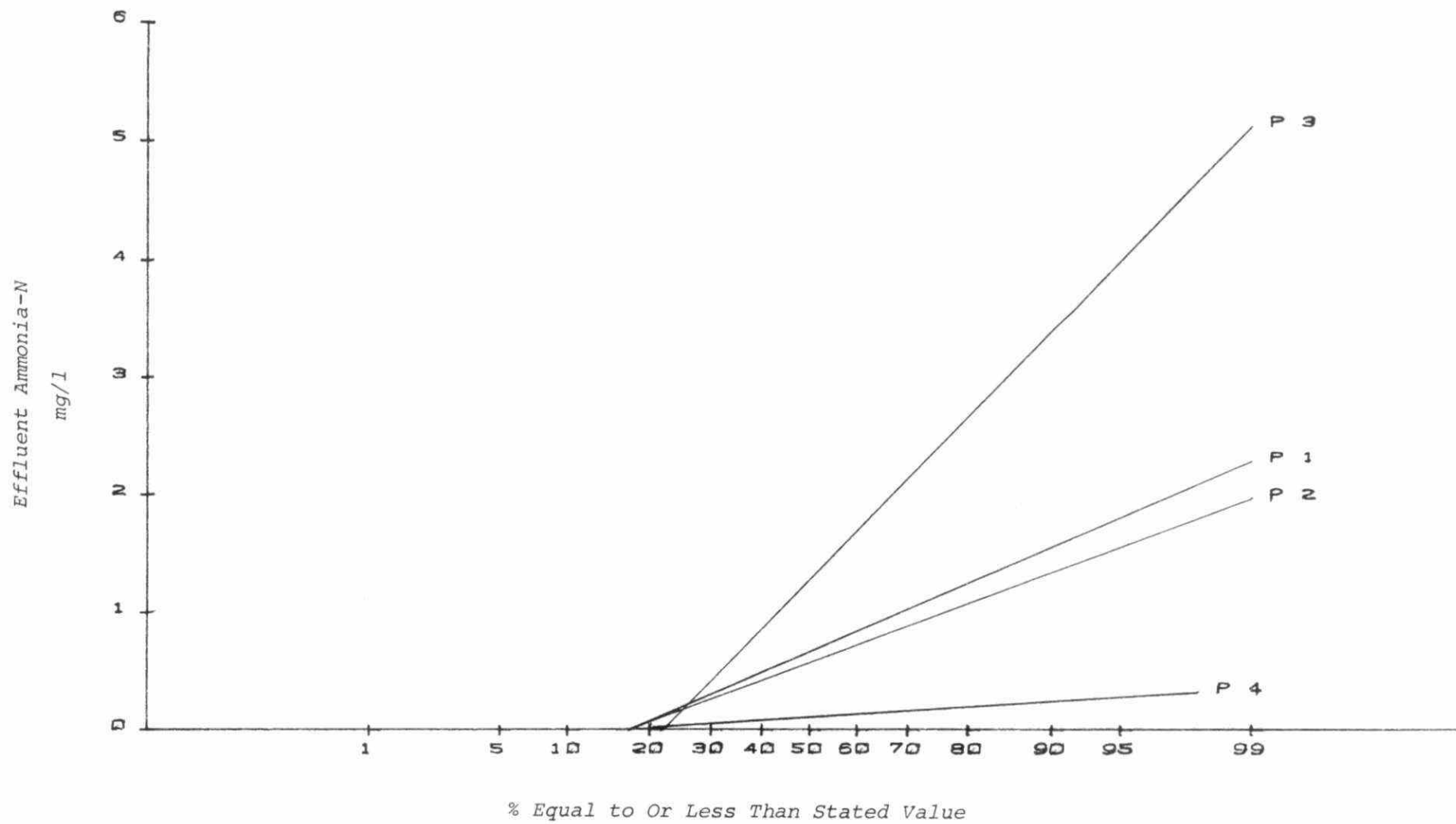


FIGURE 9. COMPARISON OF SECONDARY EFFLUENT AMMONIA CONCENTRATIONS PROGRAMS 1 TO 4

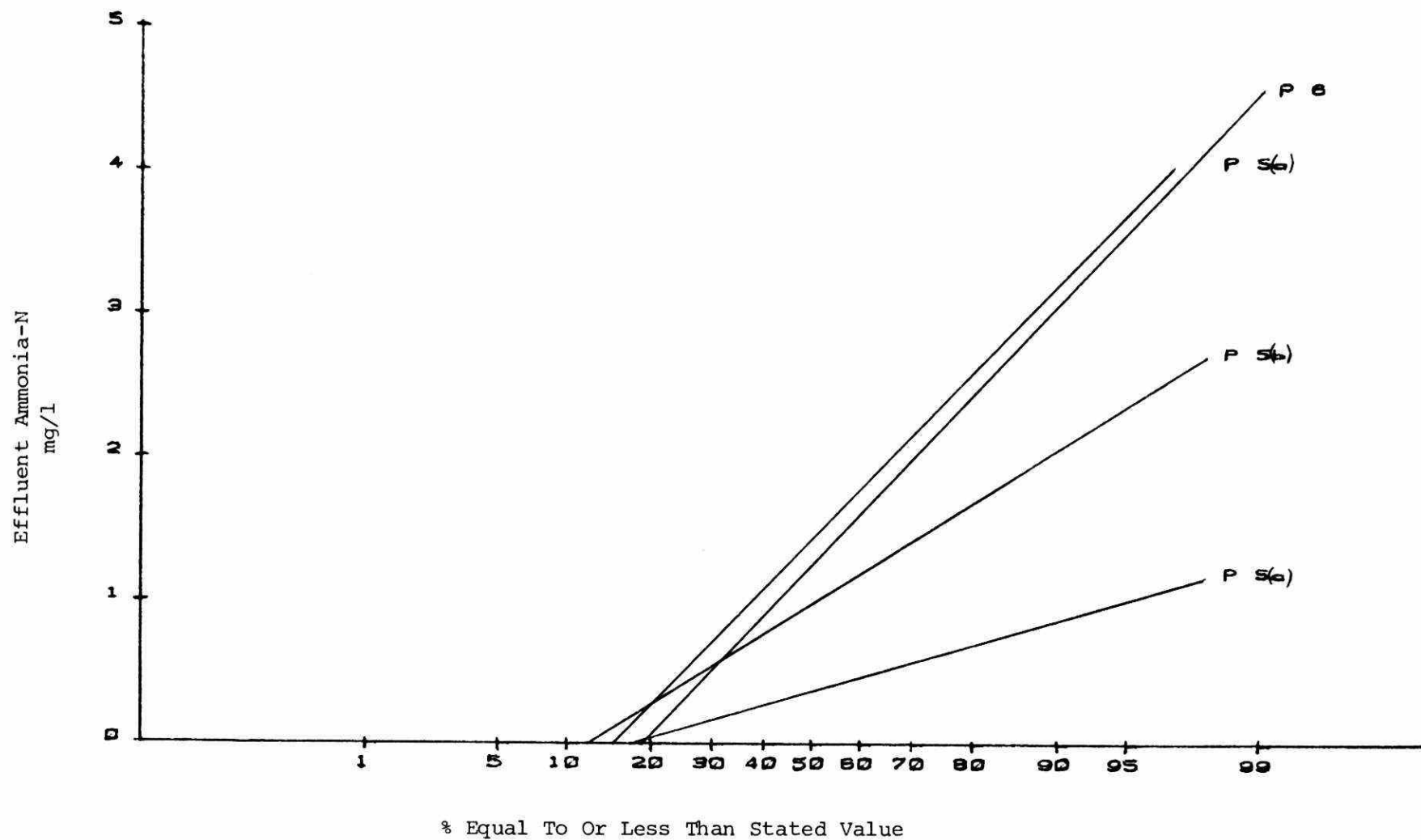


FIGURE 10. COMPARISON OF SECONDARY EFFLUENT AMMONIA CONCENTRATIONS, PROGRAMS 5 TO 6

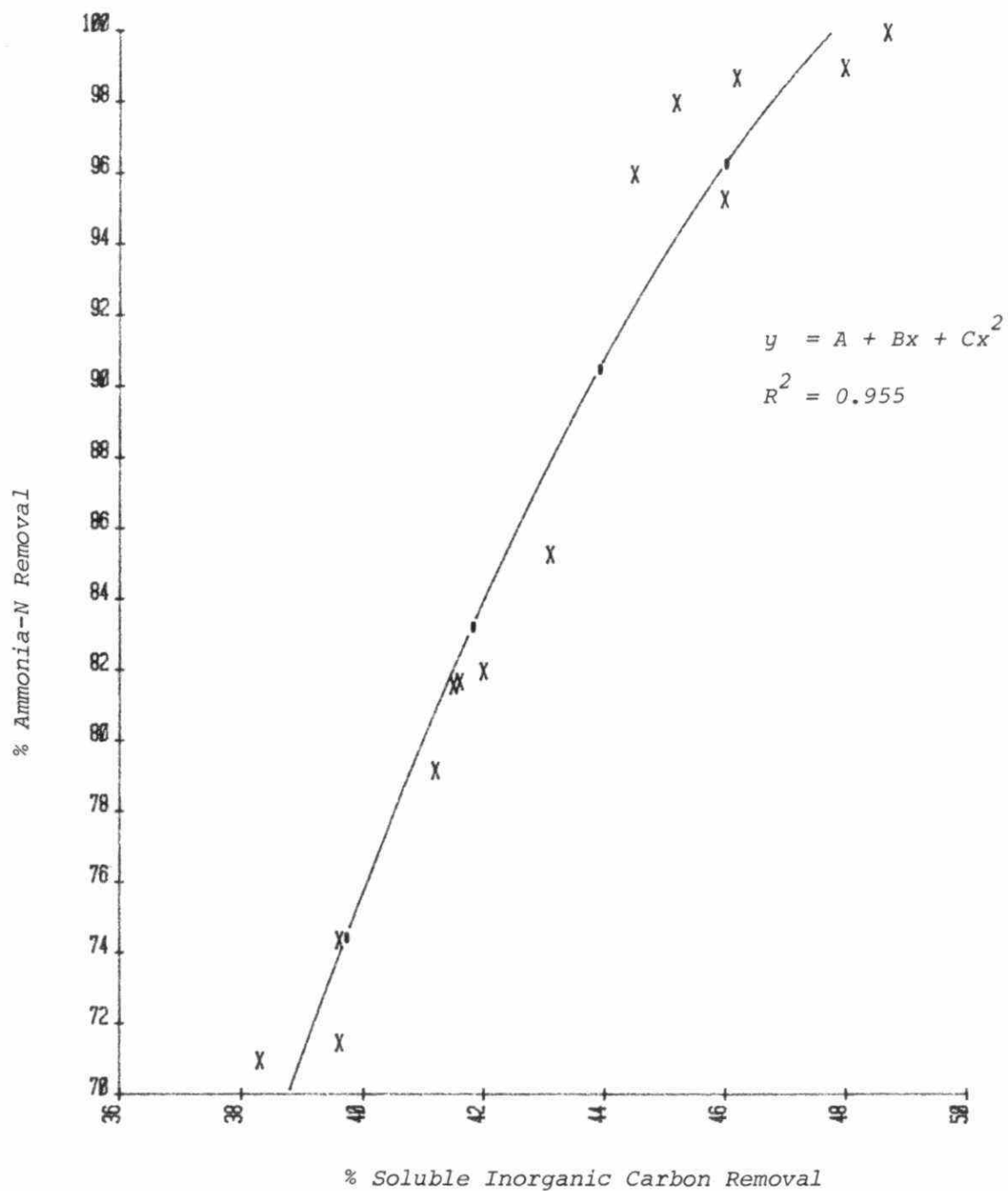


FIGURE 11. AMMONIA REMOVAL VS. SOLUBLE INORGANIC CARBON REMOVAL, PROGRAM 1

TABLE 7. EFFLUENT ANALYTICAL DATA AND PROCESS REMOVALS  
(Mean data in mg/L)

Program	Secondary Clarifier							Filtrate					
	BOD	SS	Tot. P	Tot. TKN	COD	TN	TOD	Sol. P	Alk.	NH <sub>3</sub>	NO <sub>2</sub>	NO <sub>3</sub>	TC
1	7.0	18	5.8	3.1	56	4.6	35	3.7	158	0.6	0.18	1.3	53
2	9.6	20	4.6	2.8	51	4.1	42	3.7	138	0.6	0.05	1.2	50
3	7.8	10	3.5	2.3	42	3.5	34	3.2	134	0.8	0.06	1.1	50
4	3.5	8.3	3.2	1.5	31	2.7	17	3.2	181	0.1	0.04	1.2	53
5(a)	10	13	4.2	2.8	51	3.9	43	3.6	176	0.3	0.06	1.0	50
5(b)	5.7	11	4.2	3.3	40	6.0	32	4.1	155	1.5	0.1	2.7	46
5(c)	22	32	3.5	5.7	58	9.8	92	3.5	153	1.4	0.15	3.9	47
6	14	31	1.3	4.0	53	5.4	60	0.66	143	1.2	0.04	1.4	44

Total % Removal

	1	2	3	4	5(a)	5(b)	5(c)	6
BOD	97	95	95	98	96	98	91	95
COD	91	92	93	94	93	94	92	91
SS	92	94	96	97	97	98	94	94
TN	88	90	89	90	89	84	75	87
Tot. P	34	50	51	51	54	53	60	86
Sol. P	5	24	9	3	5	-17	15	83

flow, producing afternoon and evening values up to 4 mg/L, while effluent ammonia levels changed inversely to the nitrate pattern (Figure 12).

The data presented in Table 4 show that most of the nitrate removal was attained within the denitrification basin. However, slight differences in nitrate concentrations occurred down-stream between the denitrification basin and the secondary clarifier effluent. This is further illustrated in Table 8 where differences in denitrification and total process percent nitrate may be seen.

### 3.3.1 Effects of SRT, GSA and detention time

Data analyses showed no correlation between nitrate removal efficiency and total-system SRT. This lack of correlation may be seen from the data presented in Table 8, when comparing the mean nitrate removal with the SRT in each of the programs. Similar lack of correlation between partial  $\tau^D$  (denitrification basin only) and nitrate removal is also apparent.

There is, however, an apparent relationship between sludge age - expressed in Table 8 as total sludge age (TSA), the combined weight of aeration, denitrification, and post aeration solids - and nitrate removal; higher nitrate removal corresponded to higher sludge age values. This phenomenon parallels the relationship between GSA and nitrification efficiency discussed in section 3.2.3.

Program 4, at a TSA of 43 days, produced nitrate concentrations consistently below 1 mg/L in the denitrification basin effluent. At the higher temperatures and lower TSA in Program 5, denitrification was reduced and nitrate-N values between 2 and 5 mg/L were common (Tables 4 and 8).

Improved denitrification was also coincidental with longer detention times. Comparing Program 4 with Program 5 at one-third the basin volume, nitrate removal efficiency dropped 20%, despite higher methanol dosing, denitrification rate and temperature. An additional drop of 10% nitrate removal efficiency was produced in Program 5(c) without methanol addition and at slightly lower temperatures (Table 8). An efficiency of 81% nitrate removal was accomplished in Program 6 with methanol addition, at approximately twice the detention time and TSA of Program 5.

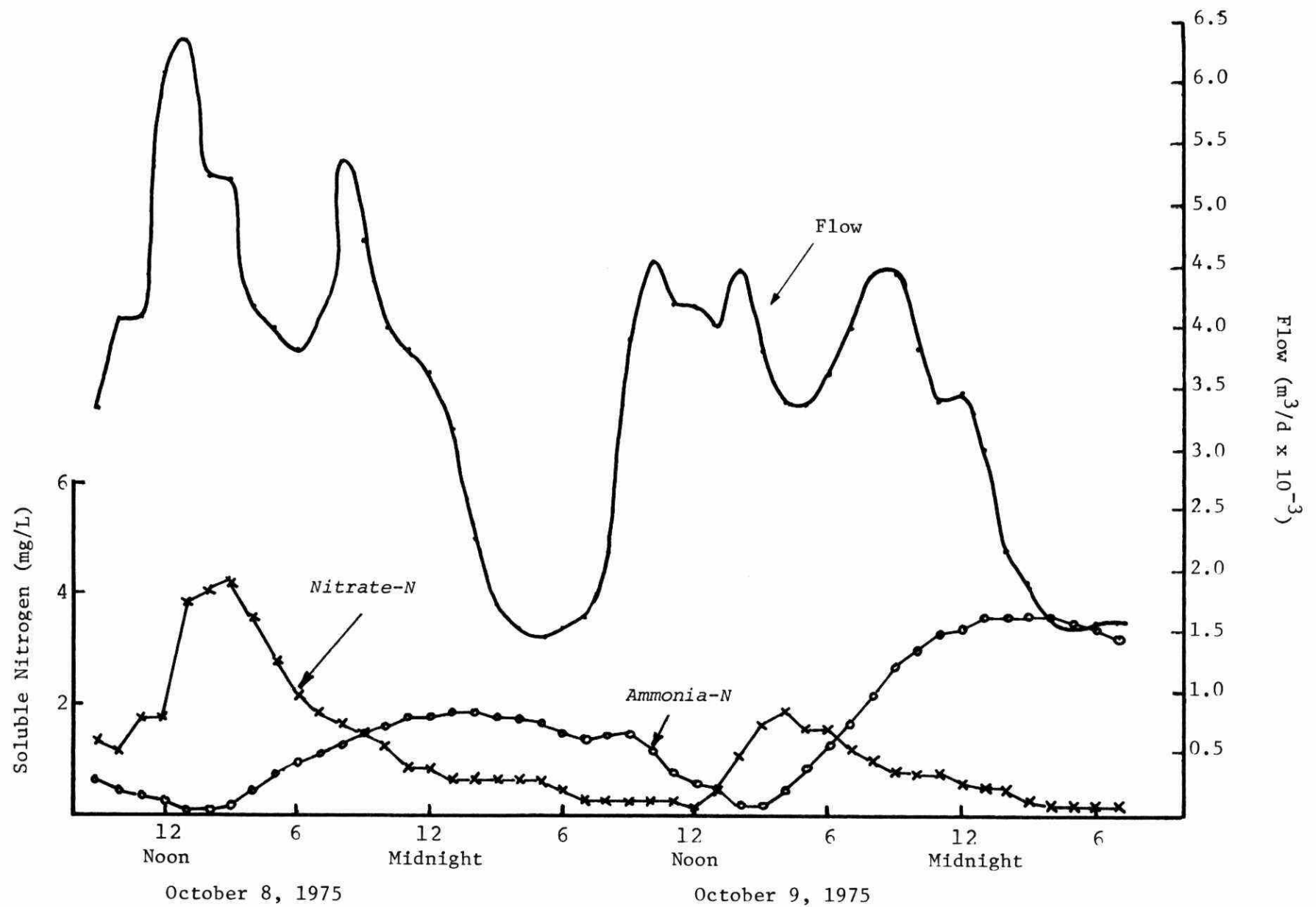


FIGURE 12. EFFECTS OF FLOW PATTERN ON EFFLUENT NITROGEN CONCENTRATION, PROGRAM 2

TABLE 8. PROCESS CONDITIONS AND DENITRIFICATION EFFICIENCY

Program	Meth/Nitrate -N Ratio	Sewage Temp. °C	Primary BOD/TKN Ratio	TSA Days	Total SRT Days	Hours Denitrification Detention		Denitrification Rate (kg nitrate removed per kg VSS/day)	% Nitrate Removed	
						Nominal	Actual*		Denit.	Total
1	0.9/1	18	3.2	29	63	7.5	2.8	0.0061	87	84
2	1.2/1	16	3.3	17	67	6.2	2.5	0.0065	77	84
3	0.9/1	10	2.9	25	222	6.2	2.6	0.0077	89	87
4	0.4/1	10	3.5	43	41	7.4	3.2	0.0069	92	88
5(a)	3/1	20	3.6	12	122	2.1	1.1	0.0209	71	90
(b)	6/1	22	3.6	15	166	2.4	1.1	0.0165	73	65
(c)	nil	18	4.4	13	54	2.6	1.2	0.0172	60	64
6 (Fe <sup>+3</sup> )	3.5/1	16	3.7	29	9	4.8	3.1	0.0089	81	82

\* Detention time with recycle flow.



Analyses of grouped data from Programs 4 to 6, as shown in Figure 13, revealed a relationship between denitrification solids and nitrate removal. A rise from 65 to 90% nitrate removal resulted upon raising the basin solids from 900 to 4994 kg (2000 to 11000 lbs). Further data computation including all programs showed correlation between nitrate loading rate (wt.  $\text{NO}_3^-$  applied per wt. denitrification VSS/day) and denitrification efficiency (Figure 14). Upon decreasing the loading rate from 0.025 to 0.005, a gain of 30% nitrate removal occurred.

### 3.3.2 Denitrification rate

Denitrification rate (wt.  $\text{NO}_3^-$  reduction/wt. denitrification VSS/day) averaged between 0.0061 and 0.0067 in the first four programs. Higher values (0.0172-0.0209) were observed in Programs 5(a) to 5(c) at half the denitrification basin volume (Table 8). When the basin volume was enlarged to two bays in Program 6, the rate dropped to 0.0089. Generally, this rate was inversely proportional to basin volume and weight of available denitrification solids.

The denitrification rates at OEF were lower than those cited in the literature. For example, Barnard [5] reported rates of 0.064 to 0.074, Bishop et al [4], 0.025 to 0.070, and Jank at the Canada Centre for Inland Waters (personal communication) found values of 0.036 at 10°C and 0.092 at 20°C. The results from OEF did not show any variation with temperature.

### 3.3.3 Consistency of nitrate removal

Probability plots for denitrification nitrate-N effluent levels are presented in Figures 15 and 16. For the most part, the spread of residual nitrate concentrations in the denitrification basin was narrower at lower mean values. Program 4 produced the best removals with 80% of the values being below 1.6 mg/L nitrate-N. Comparable results were observed in Programs 3 and 6 in that both produced less than 2 mg/L nitrate-N 80% of the time. Earlier programs (1 and 2), with wider fluctuations in influent flow, also produced larger variation and higher values of denitrification nitrate-N. Consistency of nitrate removal was least when the denitrification volume was decreased in Programs 5(a), (b) and (c). During this phase 80% of the nitrate-N results ranged between 3 and 7 mg/L.

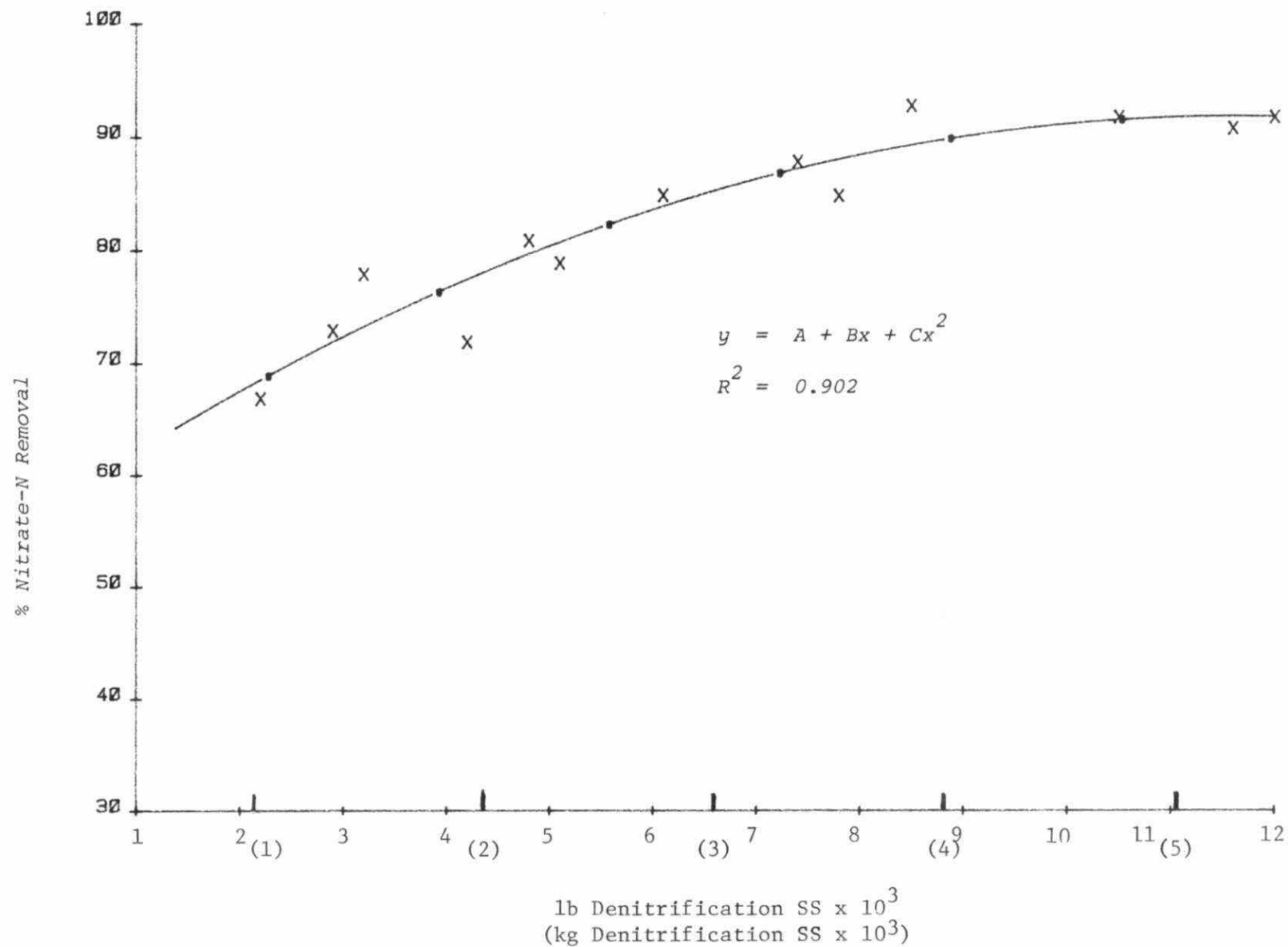


FIGURE 13. RELATIONSHIP BETWEEN NITRATE REMOVAL AND WEIGHT OF DENITRIFICATION SOLIDS, PROGRAMS 4 TO 6

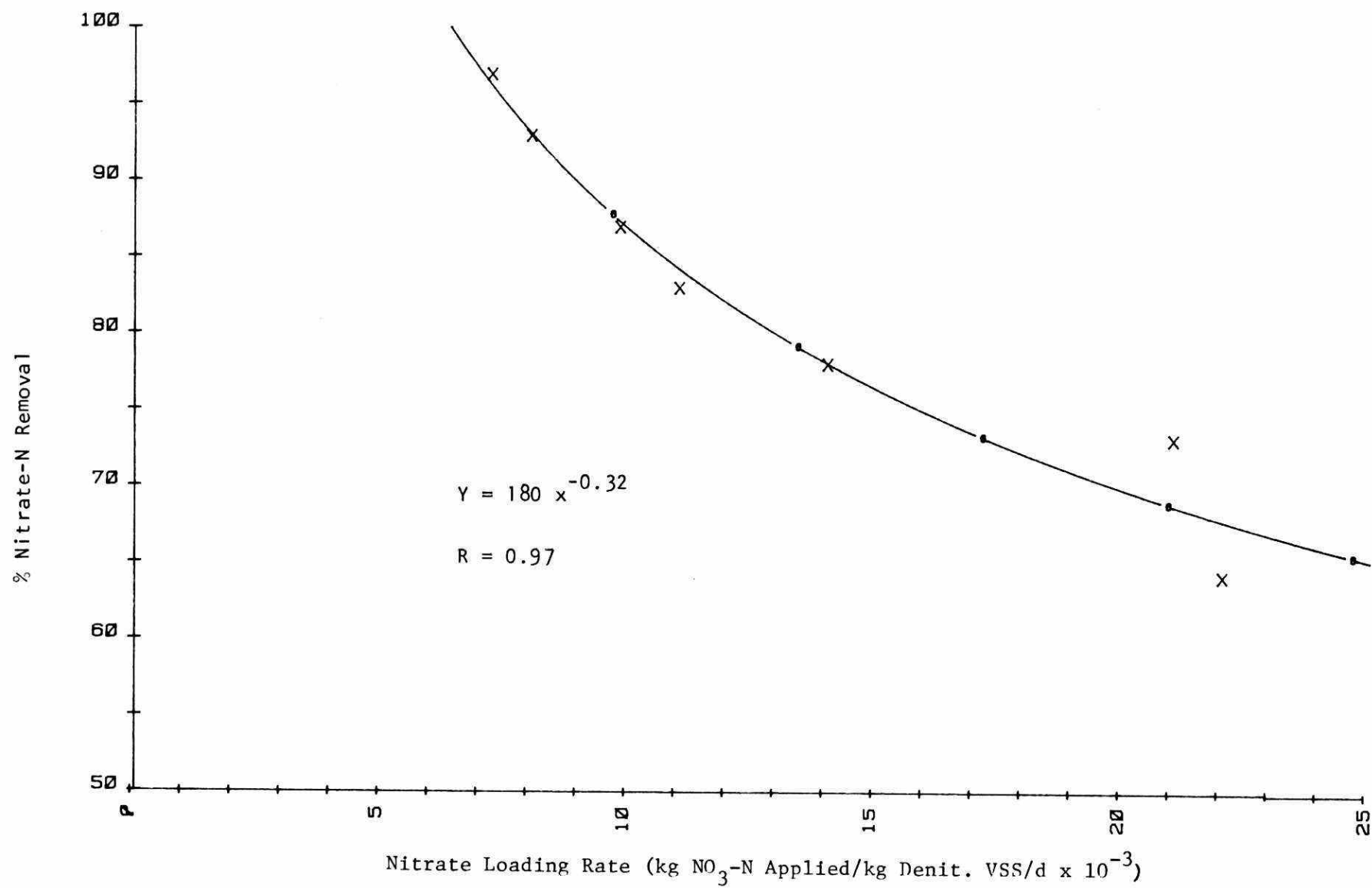


FIGURE 14. NITRATE LOADING RATE VS NITRATE REMOVAL

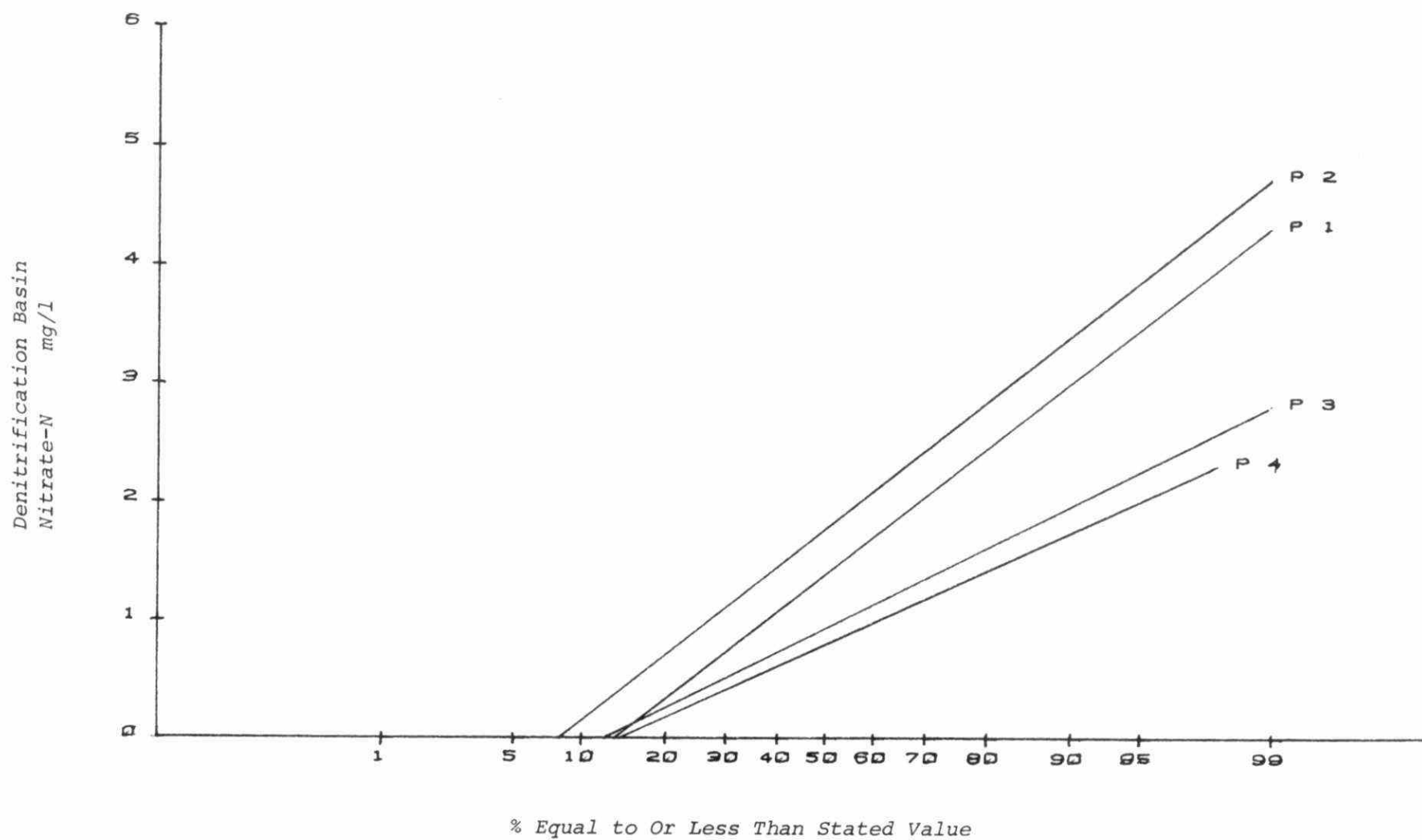


FIGURE 15. DENITRIFICATION BASIN NITRATE-N LEVELS, PROGRAMS 1 TO 4

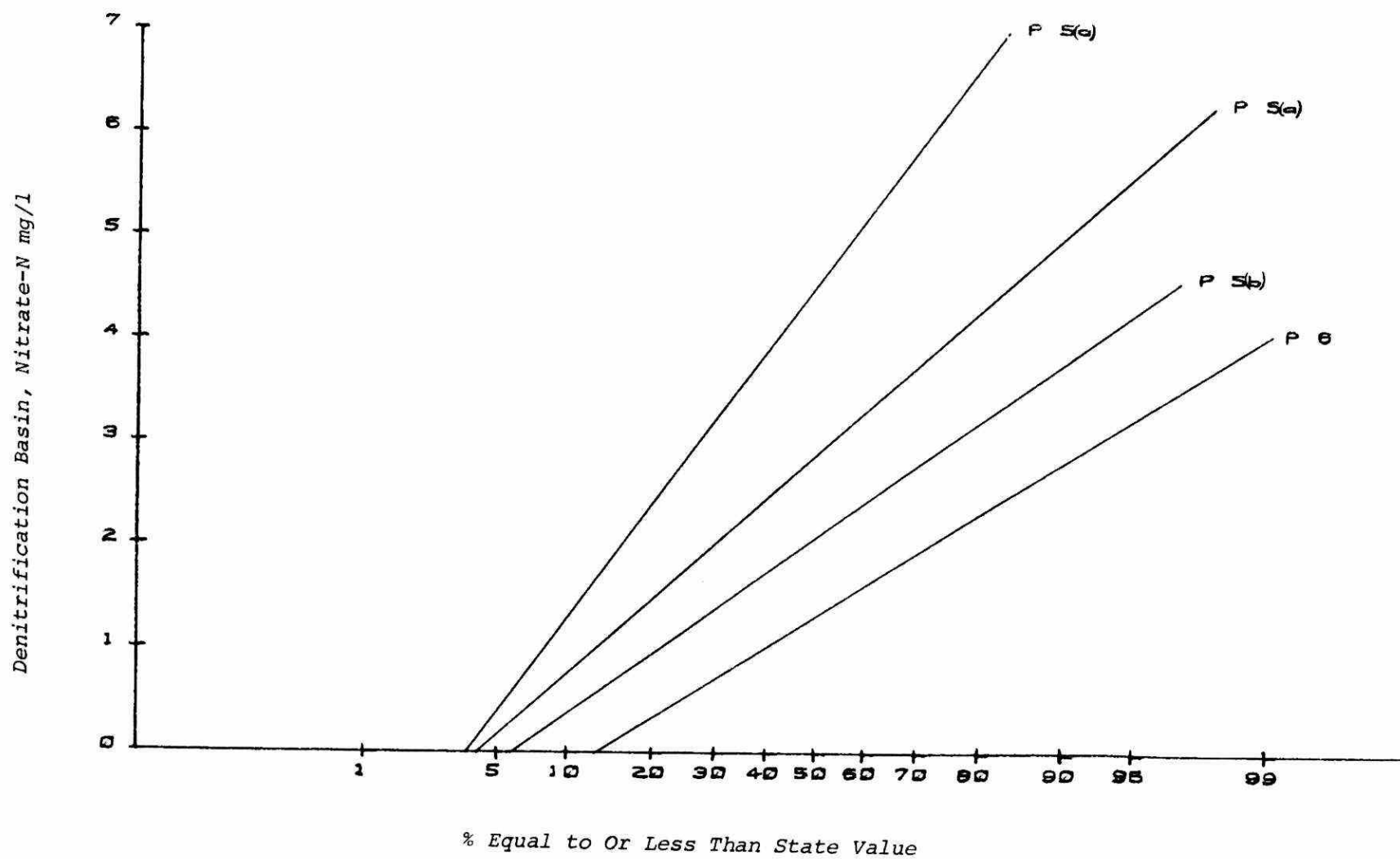


FIGURE 16. DENITRIFICATION BASIN NITRATE-N LEVELS, PROGRAMS 5 TO 6

### 3.3.4 Methanol requirements

As stated earlier in Section 2.2.1, methanol was normally fed to the denitrification basin on demand to maintain nitrate-N below 1 mg/L. The exceptions to this were in Programs 5(a) and (b) where methanol to nitrate (M/N) dosage ratios were deliberately set at 3/1 and 6/1 respectively.

A formula recommended by McCarty et al [13] was used to calculate the theoretical methanol requirement of the denitrification basin. This formula was:

$$C_m = 2.47 N_o + 1.53 N_i + 0.87 D_o$$

where:  $C_m$  = methanol dosage mg/L,  
 $N_o$  = initial nitrate-N concentration mg/L,  
 $N_i$  = initial nitrite-N concentration mg/L,  
 $D_o$  = initial DO concentration mg/L.

Using the denitrification basin influent levels of nitrate, nitrite and DO, the methanol requirement was calculated for each program (Table 9).

TABLE 9. COMPARISON OF PREDICTED AND ACTUAL METHANOL REQUIREMENTS OF THE OEF SYSTEM

Program	$C_m$ (mg/L)		Meth./Nitrate-N Ratio	
	Predicted	Actual	Predicted	Actual
1	17	7	2.2/1	0.9/1
2	22	9	3.0/1	1.2/1
3	25	8	3.1/1	0.9/1
4	31	4	3.0/1	0.4/1
5(a)	30	30	3.0/1	3/1*
5(b)	25	47	3.2/1	6/1*
5(c)	32	nil	2.8/1	nil
6**	27	28	3.3/1	3.5/1

\* Non-demand methanol addition.

\*\* Ferric chloride addition.

The demand  $C_m$  and the M/N ratio were both considerably lower than the predicted values during the first four programs. Higher methanol dosing in Programs 5(a) and (b) were equal to and twice the theoretical requirement, respectively. However, as stated earlier (section 3.3.1), higher carbon supplement during this period did not maintain denitrification nitrate-N at or near the 1 mg/L level. In Program 6, with ferric chloride addition and twice the basin volume of Program 5, the methanol requirement was close to the predicted value (Table 9).

Upon examining project data, poor correlation was noted between denitrification efficiency and M/N ratio. During the latter part of Program 4 higher denitrification solids concentrations did show, however, some influence on methanol requirements. As shown in Figure 17, raising available denitrification solids, in the time period February through April, resulted in a noticeable reduction in methanol requirement. Higher M/N ratios were required at denitrification solids of less than  $5 \times 10^3$  kg SS ( $11 \times 10^3$  lbs) to maintain 1 mg/L nitrate-N. During April, little or no methanol was used following a period when the solids weight was approximately  $5.5 \times 10^3$  kg ( $12 \times 10^3$  lbs). At this time, nitrate-N concentrations were maintained below 0.7 mg/L in the denitrification basin.

### 3.3.5 Total nitrogen removal

Total nitrogen (TN) is defined in this report as being the sum of TKN,  $\text{NO}_2^-$  and  $\text{NO}_3^-$ . The effluent values shown in the time series plot of Figure 18 are therefore a result of both biological reactions and sedimentation efficiencies.

Expanding on what was stated in Section 3.1, TKN reductions in the primary clarifier were due to suspended solids removal. These removals improved with either low solids loading or raised wastewater temperature; with temperature appearing to be the more controlling factor. Temperature is not included in Figure 18, but lower TN removals were observed in the primary clarifier during the winter months.

For the first four programs, secondary clarifier effluent TN's were consistently below 5 mg/L at influent (raw sewage) levels mostly between 25 and 50 mg/L (Figure 18 and Table 7).

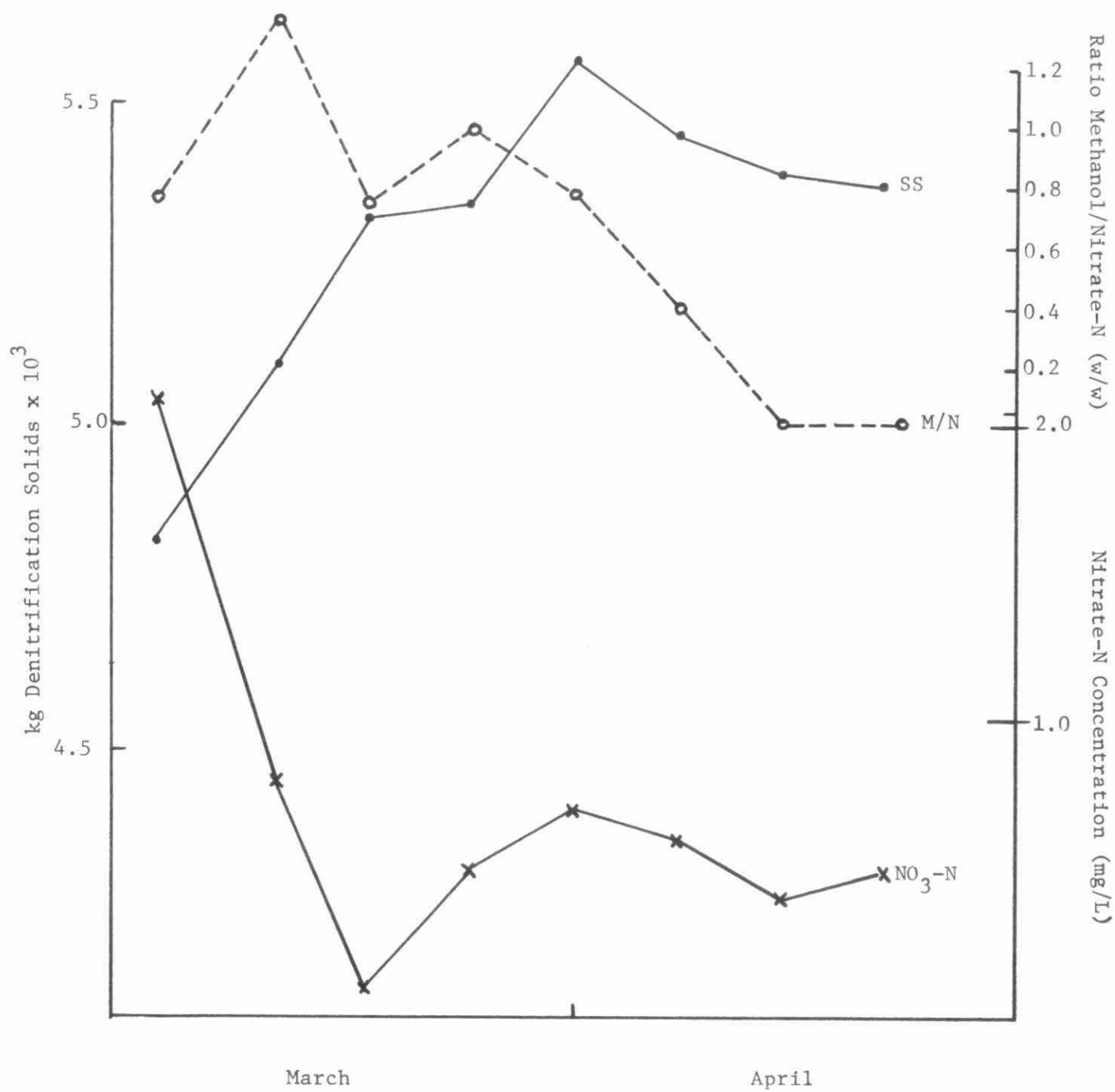


FIGURE 17. EFFECT OF DENITRIFICATION SOLIDS ON METHANOL REQUIREMENTS FOR NITRATE REMOVAL (Weekly Means)



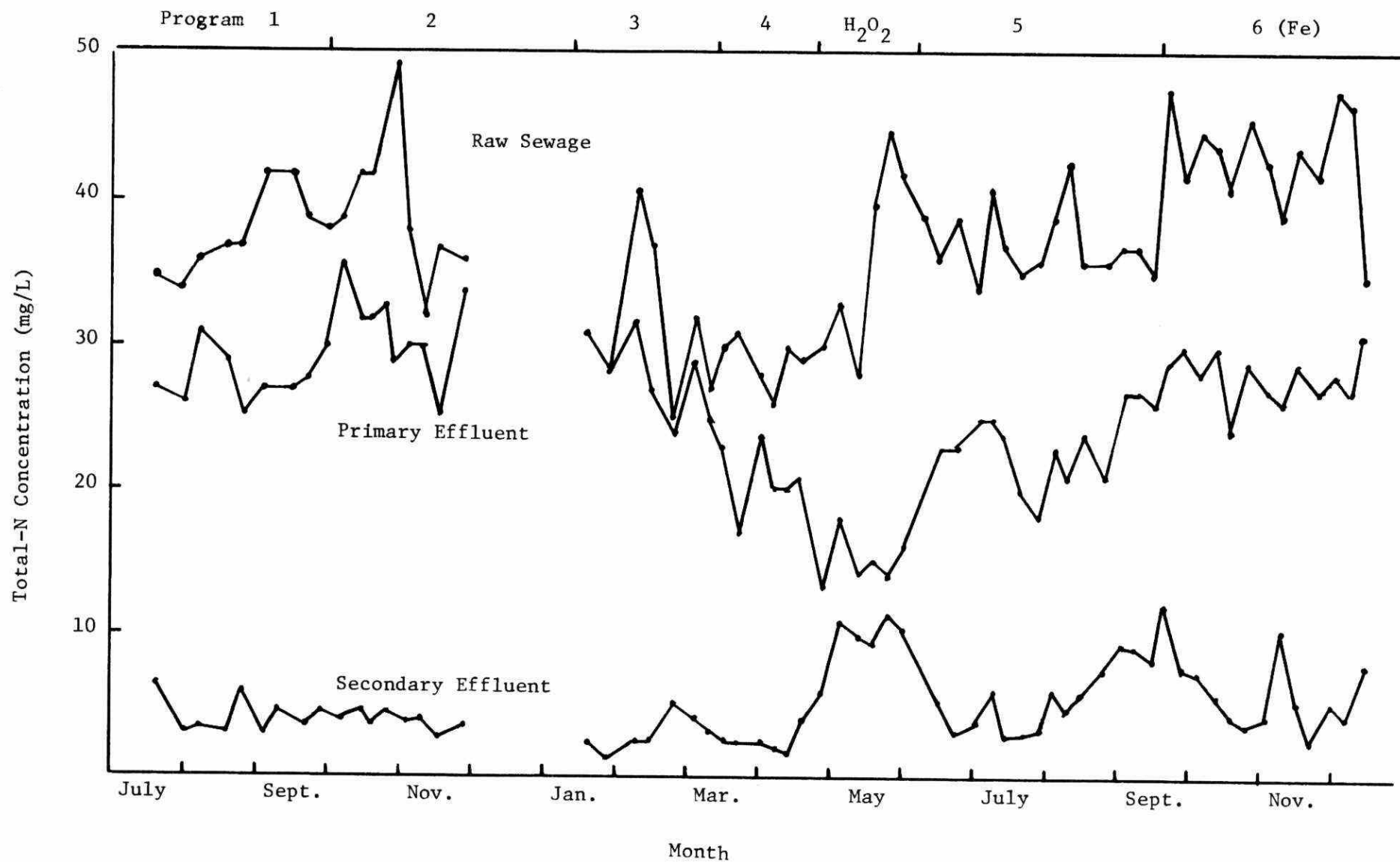


FIGURE 18. SYSTEM TOTAL-N CONCENTRATIONS (OEF)

A variation in effluent TN from 4 to 12 mg/L was produced in Program 6 and, as in Program 5(c), higher concentrations of suspended solids increased the TN. Mean program removals of TN ranged from 88 to 90% in Programs 1 to 4. As expected, a drop in TN reduction occurred in Programs 5(b) and (c) with decreased nitrate removal. An improvement to 87% TN removal was shown in Program 6 over the previous two programs (5(b) + (c)). This was mainly due to reduced effluent soluble nitrogen.

### 3.3.6 Effects on calcium alkalinity

A specialized sampling schedule over a ten-day period during Program 4 revealed noticeable calcium alkalinity changes throughout the nitrification-denitrification process (see Table 10). A reduction of 7.1 kg of calcium alkalinity occurred per kg ammonia-N removed in the aeration basin. Following this, a gain in alkalinity concentration was realized in the denitrification basin which amounted to 4.7 kg  $\text{CaCO}_3$  per kg soluble N ( $\text{NO}_2^-$  and  $\text{NO}_3^-$ -N) reduced. Mean program data as shown in Table 10, revealed alkalinity removals of between 4.2 in Program 4 to 6.6 kg alkalinity per kg soluble-N reduced in Program 6.

The theoretical exchanges in alkalinity due to nitrification and denitrification reactions are [11]:

7.2 kg of  $\text{CaCO}_3$  is removed per kg ammonia oxidized and

3.6 kg of  $\text{CaCO}_3$  is released per kg nitrate reduced.

The net alkalinity loss of a nitrification-denitrification system should then be the difference of the above two values (3.6). By comparison, the results from OEF show a mean net loss (Programs 1 to 5(c)) of 2.2, a ratio of 0.6 to the theoretical value. Similar results were reported by Sutton et al [14] where a ratio of observed to theoretical net loss of 0.8 was shown. This value was obtained from a post-denitrification single-sludge pilot process, similar in design to that at OEF.

### 3.4 Phosphorus Removal

Total phosphorus (TP) removal averaged between 34 and 60% for programs without ferric chloride addition with effluent values of between 3.2 and 5.8 mg/L soluble phosphorus (Table 7). Removals were mainly influenced by sedimentation efficiency, although some soluble phosphorus interchanges took place in the biological treatment stages. As shown in

TABLE 10. EFFECTS OF NITRIFICATION AND DENITRIFICATION ON  
CALCIUM ALKALINITY (Mean Values)

Sample	<u>Concentration mg/L</u>			Alkalinity as CaCO <sub>3</sub>
	NH <sub>4</sub> -N	NO <sub>2</sub> -N	NO <sub>3</sub> -N	
Primary Effluent	16.2	0.02	0.2	221
Aeration	2.9	0.35	5.1	126
Denitrification	1.1	0.10	0.6	151
Secondary Effluent	0.7	0.05	0.5	150

Reductions

Program	kg (lbs) Sol.-N/day		kg Alk./kg Sol.-N
1	41	( 91)	4.3
2	61	(135)	5.6
3	62	(136)	5.3
4	38	( 84)	4.2
5(a)	53	(116)	4.3
5(b)	32	( 71)	4.8
5(c)	39	( 86)	5.2
6*	51	(112)	6.6

\*Ferric chloride addition.

Table 11, a mean uptake of 0.8 mg/L of soluble P was experienced in the aeration basin followed by a release of 1 mg/L in the denitrification basin. Soluble P values were slightly lower in the secondary effluent, but resembled levels in the primary effluent.

With ferric chloride addition to the post-aeration basin (in Program 6) soluble P was reduced considerably in the secondary stage. Furthermore, the phosphorus release previously noted in the denitrification basin was not detected (Table 11).

TABLE 11. SOLUBLE P DISTRIBUTION IN THE OEF N-REMOVAL PROCESS (mg/L)

	Primary Effluent	Aeration	Denitrification	Secondary Effluent
Without Ferric Chloride	4.0	3.2	4.2	3.9
With ferric Chloride (mean dosage 12 mg/L $\text{Fe}^{3+}$ )	4.0	1.1	0.9	0.7

In relation to primary effluent total P, the  $\text{Fe}^{3+}$ :P ratio required to achieve a 1 mg/L effluent total phosphorus concentration appeared to be about 2.5:1 (see Figure 19). Values below 1 mg/L were achieved at ferric chloride dosages above 20 mg/L as  $\text{Fe}^{3+}$  and with  $\text{Fe}^{3+}$ /TP ratios exceeding 4:1.

#### 3.4.1 Effects of ferric chloride on secondary sedimentation

Process mixed liquor SVI dropped with the addition of ferric chloride to the post-aeration. As shown in Figure 20, aeration SVI decreased to values below 200 mL/g with increased iron addition. Denitrification or post-aeration SVI were not included in Figure 20, as meaningful one-half hour settlings could not be attained due to high ML solids concentration.

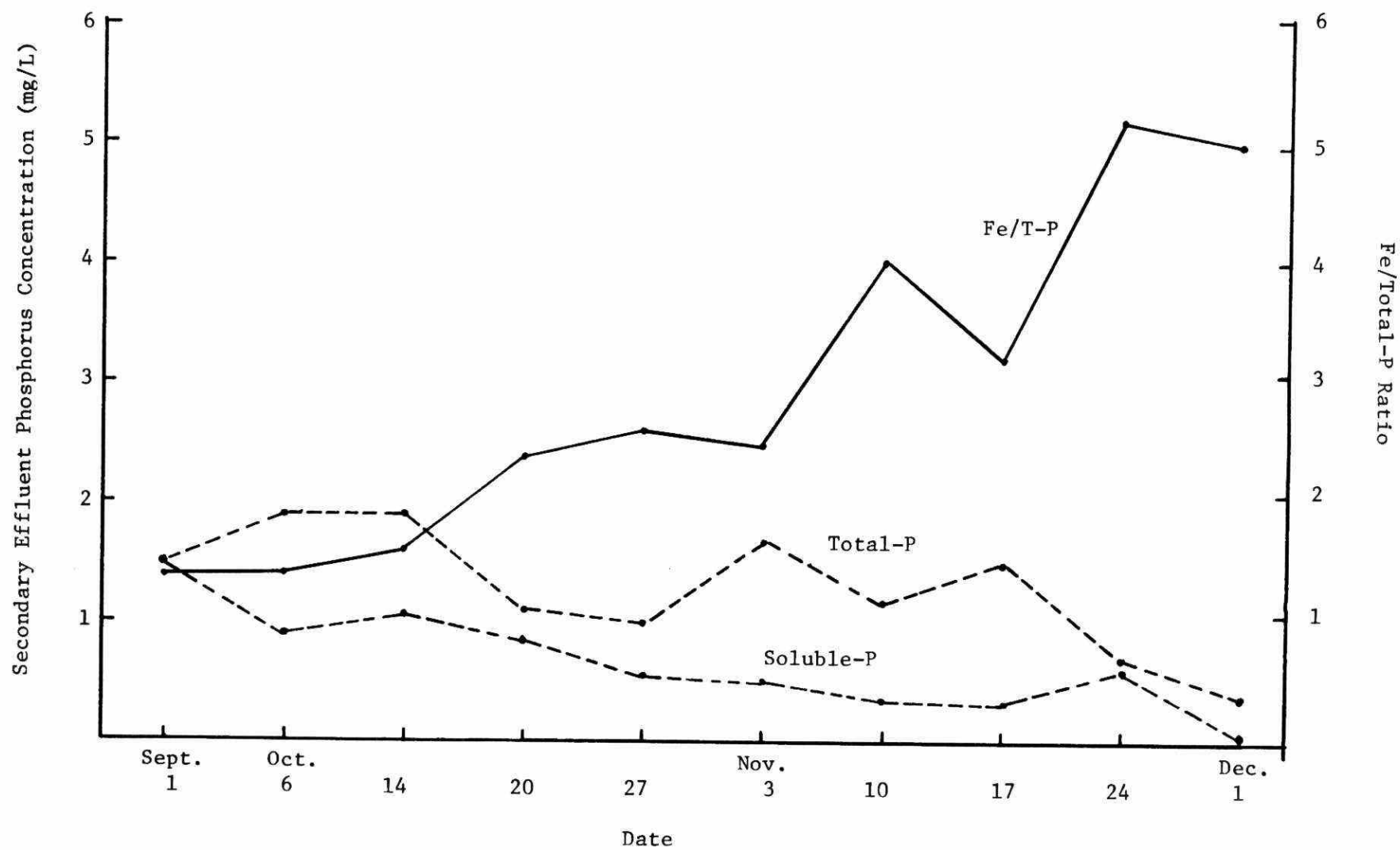


FIGURE 19. INFLUENCE OF Fe TO TOTAL P RATIO ON EFFLUENT PHOSPHORUS LEVEL (Weekly Means)

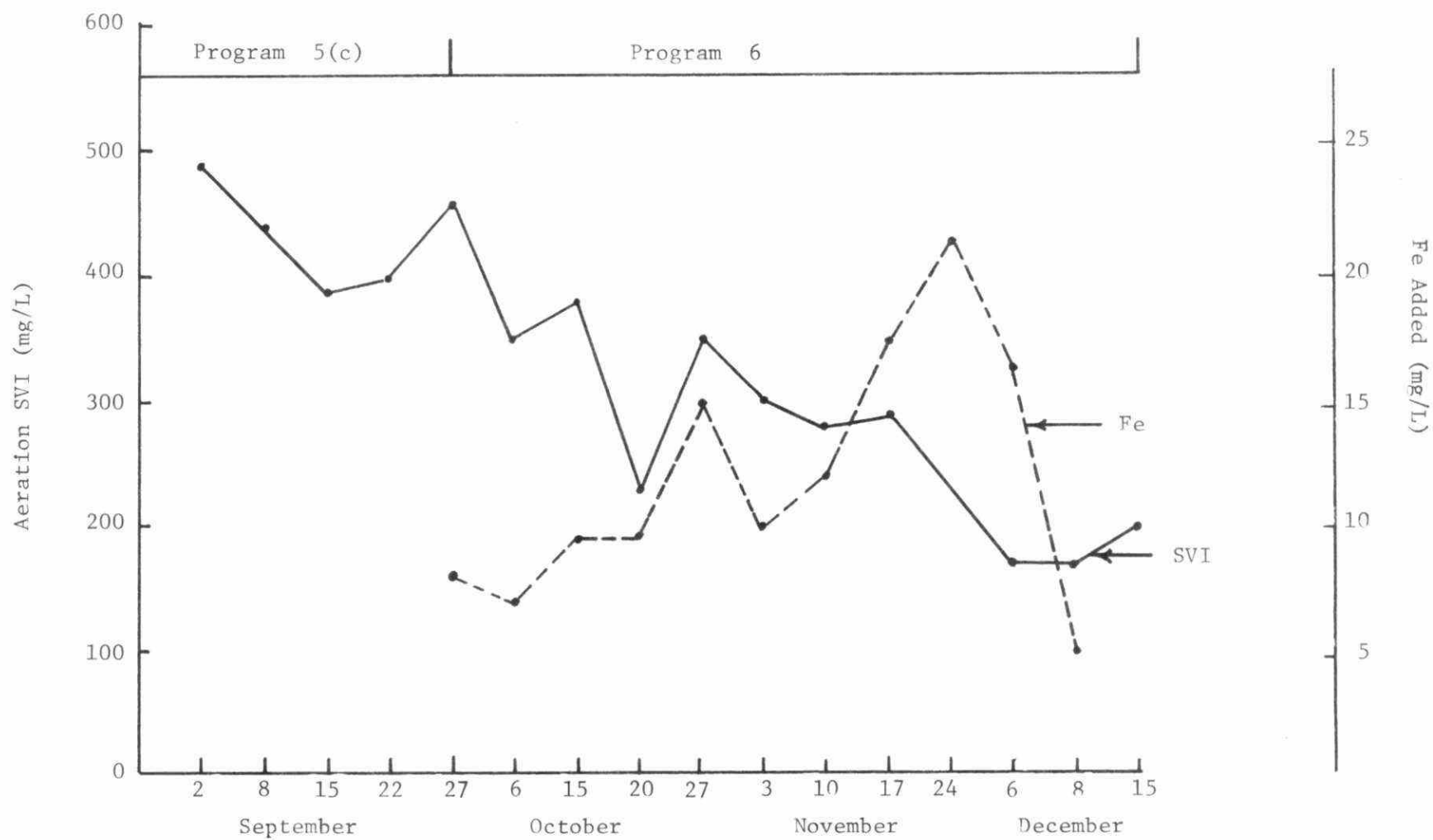


FIGURE 20. CHANGE IN AERATION SVI WITH FERRIC CHLORIDE ADDITION TO OEF PROCESS (Weekly Means)

As previously outlined in Section 2.4, the point of clarifier blanket loss was determined during various phases of the project. Two 18.3 m (60 ft) diameter clarifiers showed a total allowable solids loading of approximately 1.5 kg/m<sup>2</sup>/h (0.3 lbs/ft<sup>2</sup>/hr) at an SVI above 400 mL/g before the point clarifier sludge blanket loss occurred. The same process with ferric chloride addition and resultant lower SVI required, and operated with one clarifier, showed an allowable solids loading of 3.4 kg/m<sup>2</sup>/h (0.3 lbs/ft<sup>2</sup>/hr). Slightly higher solids loading rates were derived following the study, using the facility as a nitrification plant only. Allowable overflow rates followed a similar pattern to solids loading; increased values were obtained with decreased SVI resulting from the addition of ferric chloride (see Table 12).

TABLE 12. SOLIDS AND HYDRAULIC LOADING AT POINT OF SECONDARY CLARIFIER BLANKET LOSS

Program	Aeration SVI	Solids Loading* kg SS/m <sup>2</sup> /h (lbs SS/ft <sup>2</sup> /hr)	Clarifier Overflow m <sup>3</sup> /m <sup>2</sup> /d (lg/ft <sup>2</sup> /day)
N-Den. Without Iron	460	1.56 (.32)	9.3 (191)
N-Den. With Iron	198	3.61 (.74)	17.5 (357)
Nitr. Only	200	4.20 (.86)	21.6 (442)

\*Solids loading data do not include recycle flow.

#### 3.4.2 Effects of phosphorus removal on alkalinity

As stated in Section 3.3.5, the alkalinity loss from the nitrification-denitrification process was increased in Program 6 with ferric chloride addition. A mean alkalinity removal ratio of 4.8 kg CaCO<sub>3</sub>/kg soluble-N reduced was calculated from Programs 1 through 5. Using this value as a probable baseline for Program 6, the extra alkalinity removed per kg Fe<sup>3+</sup> added would be 1.8 kg (6.6 - 4.8) (see Table 10).

### 3.5 Treatment of High SVI

#### 3.5.1 Peroxide addition

Throughout the first four programs, a bulky mixed liquor was experienced in the OEF process. Half-hour settling tests on denitrification and post-aeration mixed liquors often were between 98 and 100% making a meaningful  $\text{cVI}$  calculation difficult. Aeration SVI's often ranged between 400-500 mL/g at a suspended solids concentration approximating 2200 mg/L.

Microscopic observations consistently revealed deflocculation and numerous filamentous organisms in the denitrification mixed liquor, and to a lesser extent in the aeration mixed liquor. On adding  $\text{H}_2\text{O}_2$  to the recycle sludge between Programs 4 and 5, a drop in SVI to between 200 and 300 mL/g was realized. However, a loss in system nitrification and consequently in nitrogen removal occurred. As shown in Figure 21, nitrogen removal dropped from an average of 90% (in the latter two weeks of Program 4) to approximately 65% efficiency in the first addition period ( $A_1$ ). A further drop, followed by a rise in nitrogen removal, resulted in recovery period ( $R_1$ ) with a return of high SVI. A higher dosage of  $\text{H}_2\text{O}_2$  at 127 mg/L to the return sludge over 24 hours ( $A_2$ ) once again reduced SVI but caused a further loss of nitrogen removal (Figure 21). The addition of  $\text{H}_2\text{O}_2$  was then abandoned and recovery period ( $R_2$ ) was implemented until nitrification was regained and Program 5 commenced.

#### 3.5.2 Screen-centrifuge concentrator

Short-term pilot studies with the screen-centrifuge on denitrification mixed liquor showed an improved sedimentation rate on passing through the 325-mesh screen. Microscopic observations revealed that the filamentous networks had been disrupted and floc particles were free of attached forms. An influent to effluent solids reduction of 45% resulted, with 30 to 45% of the centrifuge flow appearing as effluent; the remainder being screen concentrate. Effluent mixed liquor SVI ranged from 180 to 200 mg/L, which was approximately half of the aeration mixed liquor level. Visual observation of the mixed liquor during settling showed large dense floc particles, but a turbid supernatant remained.



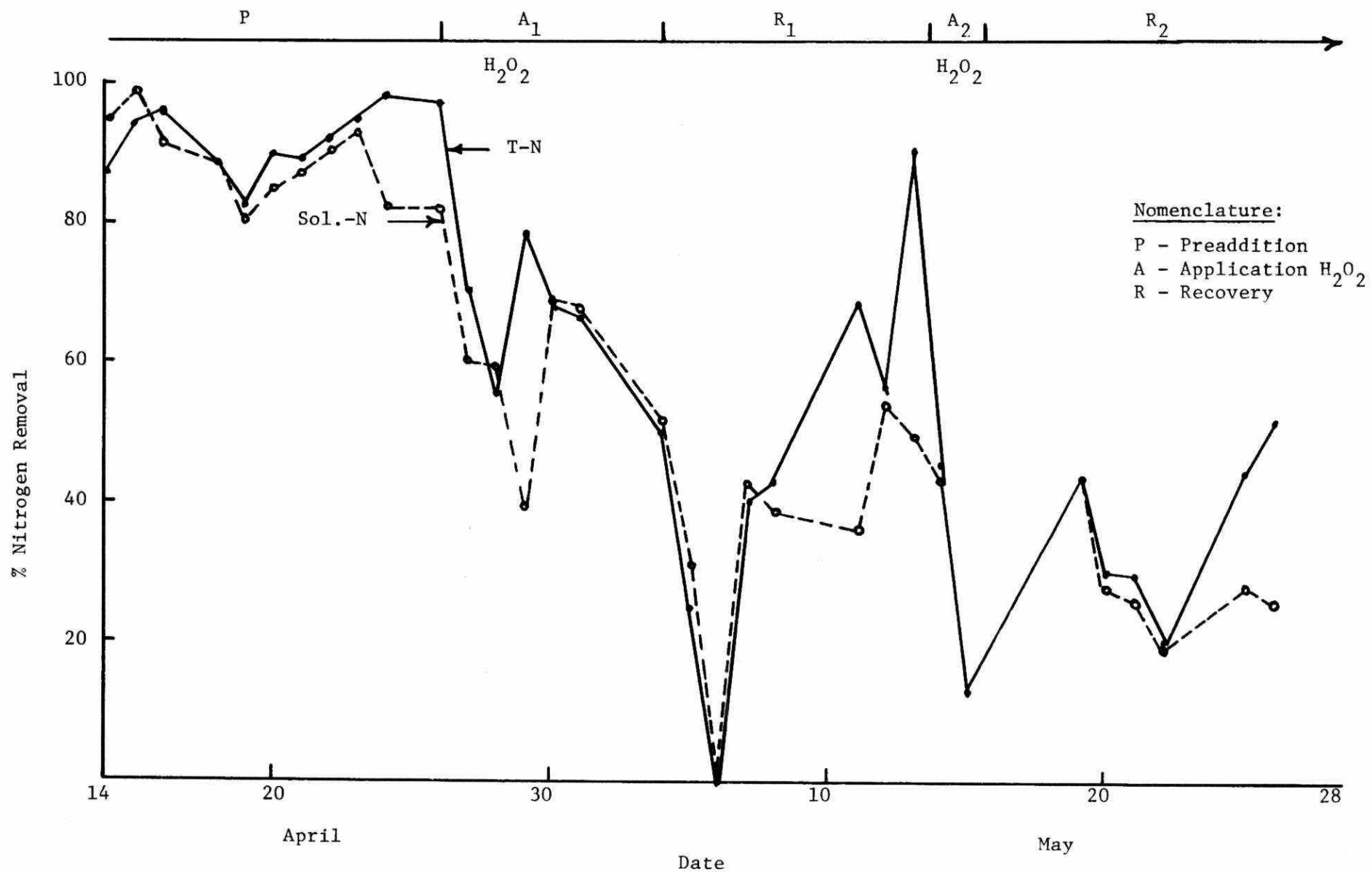


FIGURE 21. EFFECT OF HYDROGEN PEROXIDE ON OEF NITROGEN REMOVAL

Using a coarse screen (165-mesh), a higher portion appeared as effluent (80% of flow) and a 20% reduction in denitrification solids occurred. The breakup of filamentous networks and reduced SVI were not observed in these runs.

Consistent high levels of mixed liquor dissolved oxygen were measured in the screen-centrifuge effluent chamber. These values normally ranged from 5 to 7 mg/L, compared to influent concentrations of less than 0.5 mg/L for denitrification mixed liquor.

### 3.6 BOD<sub>5</sub>, COD, TC, and TOD Removal

Effluent BOD<sub>5</sub> levels were mainly influenced by SS concentrations. Mean values of between 3.5 and 22 mg/L BOD<sub>5</sub> were produced during the programs (Table 7). The soluble carbonaceous components were relatively unchanged following aeration through to the secondary effluent. Only a slight rise in TC levels was seen in the denitrification basin, with the most notable being observed in Program 5(a) with a rise of 11 mg/L over aeration levels. TC values in the secondary effluent for the most part reflected those of the denitrification basin (Tables 2, 4, and 7). Further data analyses showed no correlation between dissolved TC and nitrogen removal within the ranges produced in the OEF system.

Consistent COD removals above 90% were attained across the total process. Effluent values, ranging between 31 and 58 mg/L, rose and fell with suspended solids concentrations.

Tests by the MOE Central Laboratory on the OEF secondary clarifier effluent indicated the ratio (or R factor) of BOD<sub>20</sub> to BOD<sub>5</sub> to be 3. Calculated means for all programs are presented in Table 6 and show an effluent TOD range of 17 to 92 mg/L for Programs 4 and 5(c), respectively. Weekly means for effluent TOD and BOD<sub>5</sub> for Programs 1 to 4 are presented in Figure 22.

### 3.7 System Solids Yield

#### 3.7.1 Programs 1 to 4

An extremely low solids growth was observed in the nitrification-denitrification system. Wasting of recycle mixed liquor was only implemented when denitrification solids rose above 5500 mg/L; consequently

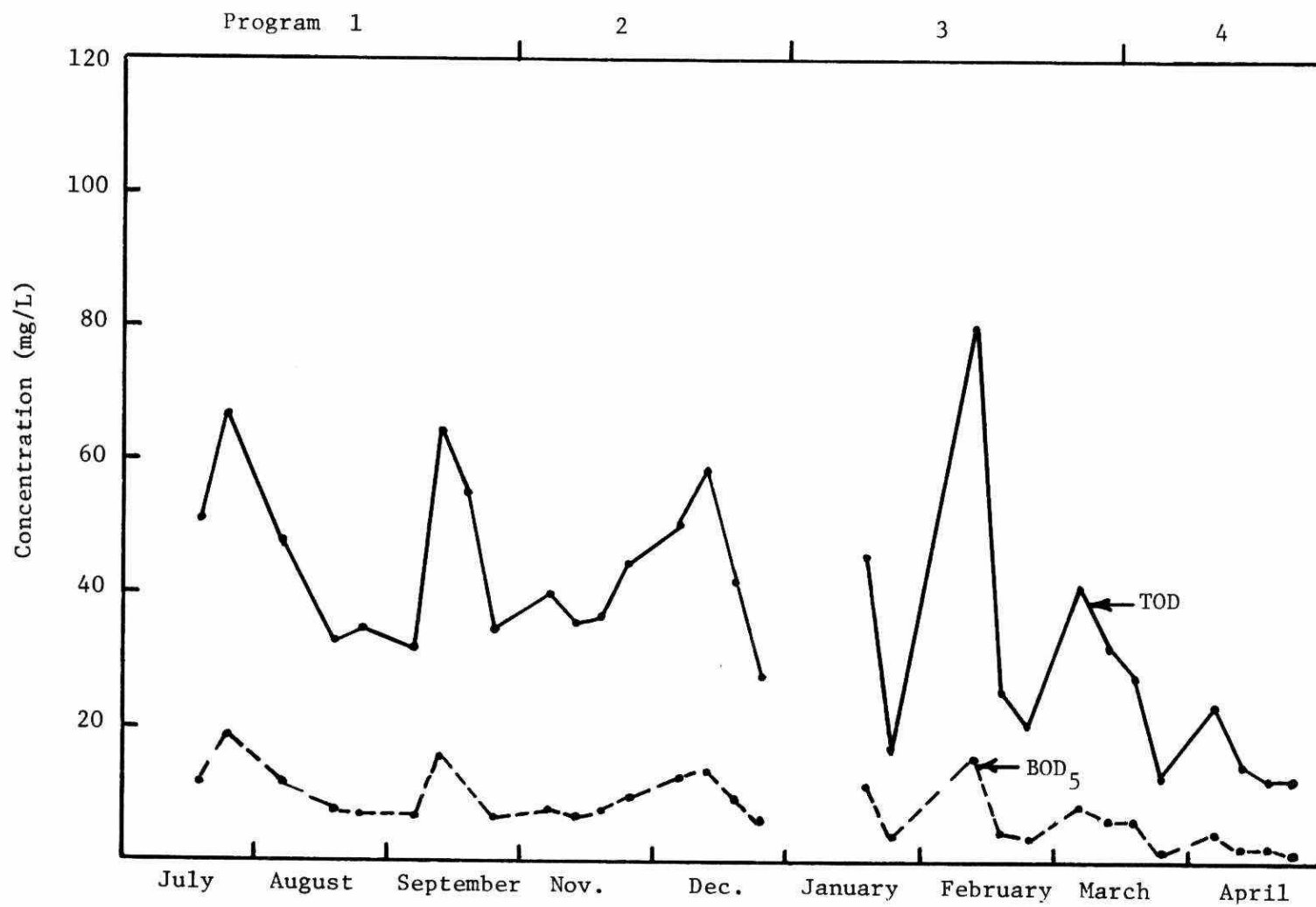


FIGURE 22. OEF NITRIFICATION-DENITRIFICATION PROCESS EFFLUENT TOD AND BOD

only occasional batch wasting was practised. Because of this, a meaningful daily sludge yield, based on BOD<sub>5</sub> utilized, was difficult to obtain.

However, a summary of solids inventory is shown in Table 13, in which process solids loss is shown with process loading. The total solids loss is defined as the weight of wasted plus secondary effluent SS per day. Programs 1 and 2 showed the highest effluent solids loss due to occasional clarifier bulking. Taking into account applied constituents (SS and BOD<sub>5</sub>) and secondary solids, Program 4 appears to have had the greatest yield of the non-P removal programs. Program 3 produced the least amount of sludge; no wasting was required and effluent solids loss was low.

### 3.7.2 Programs 5 to 6

A much lower sludge yield was shown in Programs 5(a), (b), and (c), which employed enlarged aeration and reduced denitrification capacities. No wasting was required during these programs, and solids loss to the effluent was minimal, particularly in Program 5(b) at the lowest solids loading (Table 13).

A marked increase in sludge yield was noticed with the addition of ferric chloride for phosphorus removal in Program 6. Wasting requirements were almost twice those in Program 4 without ferric addition. Most of the solids yield appeared late in Program 6 at an iron dosage exceeding 12 mg/L as Fe<sup>3+</sup>.

Estimated solids yields for Programs 4 and 6 were 0.13 and 0.55 kg SS/kg BOD<sub>5</sub> reduction/day, respectively.

TABLE 13. NITRIFICATION-DENITRIFICATION PROCESS LOADING AND SOLIDS LOSS

(Mean Values)

Program	Temp °C	Duration days	kg (lbs) Applied	BOD <sub>5</sub> /day Reduced	kg (lbs) In	Solids/day Out	kg (lbs) Solids Wasted/day	kg (lbs) Total Solids Loss/day
1	18	74	281 (618)	197 (433)	265 (584)	206 (453)	68 (149)	273 (602)
2	16	54	384 (845)	224 (494)	323 (711)	648 (1428)	127 (280)	75 (1708)*
3	10	47	304 (670)	276 (608)	220 (485)	37 (82)	nil	37 (82)
4	10	37	227 (500)	213 (470)	187 (412)	25 (55)	191 (421)	216 (476)
5(a)	20	37	348 (766)	294 (647)	293 (646)	49 (108)	nil	49 (108)
5(b)	22	33	283 (623)	266 (585)	252 (555)	35 (77)	nil	35 (77)
5(c)	18	25	367 (809)	309 (681)	258 (568)	95 (209)	nil	95 (209)
6	16	82	329 (725)	251 (552)	271 (597)	99 (219)	353 (777)	452 (996)

\* High value due to solids lost to effluent.

#### 4. GENERAL DISCUSSION

The above-normal removal of carbonaceous and nitrogenous material which occurred throughout the study in the OEF primary clarifier shows that underloading of this stage was beneficial. Higher removals reduced the loading to the secondary clarifier, thereby lowering aeration requirements for oxidation reactions. In this respect over-sized primary clarification offers benefits that are worth considering as part of the overall nitrogen removal process.

The loss of nitrification due to backmixing in the denitrification basin experienced earlier in the project suggests that a separate denitrification basin is desirable. Combined basins, such as utilized by Matsche [7] may then dictate larger air capacity and/or longer sludge age than those employed at the OEF. A series of elevated tanks offering gravity flow similar to that designed at the Canberra, Australia, nitrification plant may be more practical than the one-way gate system at the OEF [11].

Ammonia-nitrogen breakthrough at peak diurnal loading, as occurred in this study, showed that some important operation and design options should be considered. For the most part, following Programs 1 and 2, the diurnal peak-to-average flow factor of 1.5/1 was maintained. In reality, most plants would have a higher flow factor than this, which would increase the possibility of nitrogen short-circuiting to the effluent during the peak period. Larger aeration and denitrification basins at comparable flows to the OEF study may be required to counteract the effects of wide flow variation.

An option to enlarging the primary clarification or secondary stages is the application of flow equalization. Wastewater equalization prior to primary clarification, as recommended in a recent U.S. EPA manual [15], would improve primary treatment and buffer load variations to the secondary unit by reducing peak flow.

The importance of maintaining adequate aeration DO for nitrification was demonstrated early in the study. The discrepancy found (comparing nitrification rate to DO) between the OEF data and that of Nagel and Haworth (cited in [11]) was probably due to other experimental conditions.

The results support the conclusion that a minimum of 2 mg/L DO concentration is needed to maintain an adequate nitrification rate.

With nitrification proceeding at a high rate through Program 4, it was difficult to decipher whether step feeding was subsequently beneficial. The enhanced distribution of aeration DO probably helped maintain nitrification in all areas of the basin at higher temperatures in Program 5.

The relationship found between GSA and nitrification efficiency revealed a possible mode of process control. The solids level in the aeration basin was controlled by varying the recycle influent rate. The aeration solids can then be set to comply with influent solids to arrive at the GSA found optimum for full nitrification.

Since the OEF system was not run on the basis of SRT, it is not surprising that a lack of correlation existed with nitrification. The sporadic solids wasting from the recycle line common to both nitrification and denitrification basins probably produced distorted aeration SRT values.

Any effects on the process due to temperature appeared to be compensated for by the system's inherent sludge age. This was particularly true of nitrification, where a GSA of approximately 10 days appeared adequate to support nitrification throughout the seasons.

The variation observed in effluent nitrate concentrations with cyclic system loading shows that the denitrification basin reactions would also benefit from equalized flow. Since these fluctuations were in phase with flow, it can be stated that the nitrate removal system is more immediately responsive than that for nitrification. Further evidence of this was seen during recovery of system reactions following  $H_2O_2$  application. Once nitrification commenced, immediate denitrification occurred.

Even though some relationship was shown between nitrate removal and TSA, this variation of the Gould sludge age would not be functional for design and operation. However, the nitrate loading ratio could be used for these purposes on a split-return system. After determining the expected nitrate loading to the denitrification basin (kg nitrate-N/day),

the weight of required solids can be calculated by choosing an optimum ratio (e.g.,  $\leq 7 \times 10^3$  kg  $\text{NO}_3^-$ -N/kg VSS/day found at the OEF). The study showed that, by maintaining a high concentration of denitrification solids (low nitrate/solids ratio), methanol requirements are reduced. This finding agrees somewhat with that concluded in the literature [3,5,7] in that an external carbon source may not be required for efficient denitrification. This will of course be dependent on target effluent nitrate concentration and seasonal requirements.

Any effects upon N-removal reactions were evidently compensated for by a high process SRT and methanol addition. In order to relate SRT to denitrification reactions with the single-sludge, split-return system, a different study approach to that outlined in this report would be required. Both the aeration and denitrification basins should have separate solids wasting to facilitate calculation of aerobic and anoxic SRT.

Regained alkalinity via denitrification reactions can be viewed as an advantage since calcium alkalinity is lost in both nitrification and phosphorus removal reactions (depending on chemical used). Implementing a nitrification-phosphorus removal system in a locality of low alkaline wastewater may require supplemental hydroxide addition to counteract system acidity. A denitrification section could, therefore, cancel the need for or reduce the required chemical addition, depending on influent levels of alkalinity. However, the study did show that the actual net alkalinity loss of the total system was less than the theoretical value.

Greater quantities of iron-salt may be needed with the nitrification-denitrification system for phosphorus removal. The 20 mg/L of  $\text{Fe}^{3+}$  required to maintain the OEF nitrification-denitrification system effluent at 1 mg/L total phosphorus is slightly higher than normal in an activated sludge process. Prolonged addition of iron salt to the system over several months may ultimately result in lower requirements as biological/chemical reactions assume a steady state.

A drawback of the single-sludge process may be the development of a high SVI mixed liquor and the consequent adverse effect on sedimentation. As shown in the clarifier bulking tests, larger than



normal clarifiers, or multiple clarification, will be required. With ferric chloride addition, and a resultant lowering of SVI, allowable overflow rates and solids loading can be raised. It is expected that with prolonged application, the required clarification will approach the normal  $29-34 \text{ m}^3/\text{m}^2/\text{d}$  ( $600-700 \text{ IG}/\text{ft}^2/\text{d}$ ) for combined biological-chemical processes (MOE requirement).

Regardless of whether one or two clarifiers are required following post-aeration, the single-sludge system at the OEF still offers advantages over a two-sludge process. One disadvantage of a two-sludge system is that placing a clarifier between the nitrification and denitrification units would require additional recycle and wasting equipment. The split-return mode offers singular recycle and wasting, thereby requiring less operator attention than a two-sludge system.

The failure of hydrogen peroxide to lower SVI without effecting a loss of nitrification means that this chemical should not be added to this system to lower SVI. In the past,  $\text{H}_2\text{O}_2$  has been effective in treating bulking sludges caused by high process loading and low DO [16], but this combination of occurrences did not exist at the OEF.

The screen-centrifuge concentrator appears promising as a reflocculation device, but may be uneconomical because of flow restrictions caused by the 325-mesh screen. A very large unit would be required to treat the total plant flow. However, since this device produces a concentrated mixed liquor and provides aeration, it could replace the post-aeration basin. The concentrate would then be recycled to the denitrification basin to provide solids thickening. The normal recycle from the secondary clarifier would be directed to the aeration basin.

Generally, the MOE effluent quality guideline of 15 mg/L for  $\text{BOD}_5$  and SS was achieved with the OEF N-removal process. More stringent effluent requirements would necessitate effluent polishing.

Higher mean levels during iron-salt addition in Program 6, were thought to be a result of overdosing late in the program. More floc was noticed in the clarifier effluent during this period.

The removal of inorganic carbon by nitrification confirms earlier work [8] that additional TIC removal is attained with ammonia oxidation. The nitrification reactions will then aid in reducing effluent carbon levels.

For the most part, full utilization of methanol was attained in the system and its addition, therefore, did not raise the effluent BOD<sub>5</sub>. Measurement of effluent BOD<sub>20</sub> for TOD calculation appeared to be reproducible as consistent results (3 x BOD<sub>5</sub>) were obtained using ATU as a nitrification suppressant. The low solids yield found in the OEF process would constitute a saving in required sludge treatment compared to a conventional activated sludge plant. The sludge production of this system without chemical addition for phosphorus removal will probably resemble that of an extended aeration facility at similar SRT.

As the results showed, application of ferric chloride significantly increased the solids yield. However, the mean program yield was influenced by higher-than-required iron-salt addition in the latter half of the program. A ferric chloride dosage constantly maintained at 10 to 15 mg/L as Fe<sup>3+</sup> over an extended period probably would produce less sludge and provide an effluent phosphorus of 1 mg/L.

In summary, the single-sludge, split-return system gave excellent nitrogen removal with minimal methanol application. Control of solids in both aeration and denitrification basins was easily accomplished by adjusting the return split. The inherent long SRT of this system probably compensated for the effects of low temperature upon nitrogen metabolism. Monitoring of the process on-site can be accomplished easily by solids tests and field kit nitrogen analyses.

## ACKNOWLEDGEMENTS

The author wishes to acknowledge the assistance of:

Facility Chief (Ontario Experimental Facility),

Mr. B.D. D'Cruze and his staff;

Engineer: Mr. M. Lywood

Technicians: Mr. B. Bradley and Ms. R. Scicluna

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**S65**  
**1979**

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wastewater using a single-  
sludge system : volume II /  
Smith, A. G.  
78913